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Minimizing the Pollution Impact of Kraft Pulping Through Oxygen Bleaching

Chesapeake Corp. of Virginia
West Point

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16. ABSTRACT

In December, 1972, The Chesapeake Corporation started up a unique three-stage oxygen bleaching system (D/COD) to produce 272 metric air dry tons per day of 88+ brightness hardwood market pulp. This system cost less than conventional chlorination and offered potential for reducing bleach effluents. Excessive fiber losses from, and stress corrosion of, the twin roll press were the most serious operating problems experienced on start-up. These problems were satisfactorily solved. Total effluent BOD_5 and color levels indicated no decrease in BOD_5 and a 36 percent decrease in color over conventional chlorination sequences. With oxygen-stage effluent recycle and complete recovery, a 40 percent reduction in BOD_5 and a 65 percent reduction in color could be realized.

Prior to and following the bleach plant start-up, certain in-process changes were made, and systems using high-purity oxygen for wastewater treatment and weak BLO were piloted and installed. Also, an on-site cryogenic air separation plant was installed to provide low-cost oxygen to these systems. The oxygen plant, waste treatment plant, and BLO systems have operated without major process problems since 1974.

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by

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory-Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report investigates the use of a three-stage oxygen bleaching sequence in the production of bleached market pulp from hardwood. The system offers promising potential for reducing the polluting effluents normally encountered during conventional chlorine bleaching of kraft pulp. It is hoped that the report will be useful to other people in the pulp and paper industry and to any agent involved in the control of this type of pollution.

PREFACE

Ideally, any series of corrective actions taken to minimize industrial polution loads to the environment should be as systematized and interlinked as the ecosystem itself. The entire picture must be viewed with regard to the nature of the resource used, the economics of converting the resource into useful products, and the possibilities for reusing or returning waste materials generated to the environment with minimum effect on local and distant ecosystems. This difficult goal is rarely achieved, since there is a lack of knowledge about the major and the less obvious ecological interactions. Though the steps reported here by no means achieve this ultimate goal, they are systematized, somewhat by circumstance, but principally by design.

Chesapeake tries to use available natural resources wisely and actively replenishes its renewable resources by reforesting cut acreage with appropriate tree species. Within its resource procurement area (encompassing parts of Virginia, Maryland, Delaware, and North Carolina), Chesapeake finds about 50 percent hardwoods among the standing timber. These species, being short-fibered, cannot provide the quality demanded by Chesapeake's markets for its major products—namely, unbleached linerboard and paper. Since the hardwoods could not be used in proportion to their stand, they often were wasted as the more suitable long-fibered softwoods were harvested.

To reduce such waste of valuable, naturally renewable resources, Chesapeake examined several types of marketable products that would utilize more hardwood. After choosing bleached market pulp, Chesapeake did in-house research and contracted research elsewhere domestically and abroad on a variety of bleaching sequences. This research did not fall within the scope of the EPA demonstration grant. However, Chesapeake's choice of a three-stage oxygen bleaching sequence, chlorine dioxide/chlorine-oxygen-chlorine dioxide (D/COD) as the first such commercial installation, stimulated worldwide interest and drew hundreds of visitors from many countries when the plant started operation.

Claims that this bleaching sequence could reduce the pollution load of typical chlorination (specifically BOD and color) coincided with EPA's interest in process technology of low pollution potential. Thereafter, an EPA demonstration grant was discussed and negotiated with the principal focus on the oxygen bleaching systems.

Concurrent Chesapeake plans to (a) reduce pollution load in-plant, (b) revise water recycle systems, (c) treat wastewater and oxidize weak black liquor with high-purity oxygen, and (d) install an on-site air separation plant to provide the oxygen for all three processes became a peripheral part of the demonstration grant.

The total estimated cost of the systems installed, or to be installed, was $12.6 million in January 1972. The EPA grant, effective June 19, 1972 to July 1975 (with an extension to November 1975 with no additional funds) was $438,939. At the time of the grant, money available from EPA was limited. Therefore, though seemingly small as a percentage of the total, the grant amount is generous and indicative of EPA's strong commitment to encouraging the development of minimum-pollution processes.

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ABSTRACT

To improve the utilization of valuable hardwoods in timber stands used for the production of linerboard and paper, the Chesapeake Corporation produced bleached market pulp from these trees using a three-stage bleaching sequence that could reduce the pollution load of typical chlorination sequences. The oxygen bleaching sequence that was chosen produced 272 metric air dry tons per day of 88+ brightness market hardwood pulp. In addition, the three-stage sequence—chlorine dioxide/chlorine-oxygen-chlorine dioxide (D/COD)—cost less than conventional chlorination, offered promising potential for reducing polluting effluents, and could be installed with no increase in fresh water use. The sequence design, projected costs, and operation were based on purchased liquid oxygen.

Within the 16 months preceding and a like period following the bleach plant startup, certain in-process changes were made, and systems were piloted and installed using high-purity oxygen for wastewater treatment and weak black liquor oxidation (BLO). Because these systems required low-cost oxygen, Chesapeake bought and installed a cryogenic air-separation plant onsite.

The bleach plant started up in December 1972 and again in February 1973 after roll covers on the twin roll press were replaced. Although numerous other mechanical and process problems occurred during the grant period, most operating problems centered on the twin roll press. Fiber losses from the press required installation of a saveall. Stress corrosion of roll covers at the press lead to the need for a D/C stage washer ahead of the press. Also, scaling by barium sulfate and calcium oxalate partially blinded holes in the roll covers. Without oxygen-stage recycle, final BOD_5 and color measurements indicated no decrease in BOD_5 and a 36-percent decrease in color over conventional chlorination sequences. With oxygen-stage effluent recycle and complete recovery, a 40-percent reduction in BOD_5 and a 65-percent reduction in color could be realized. Oxygen-stage recycle to recovery would elevate white liquor chloride levels from 6 to 17 g/L in a kraft mill producing 20 to 25 percent bleached hardwood pulp.

The 91 metric ton/day oxygen plant operated well after May 5, 1974. No major problems occurred until after the grant period, when the molecular sieve system for removing carbon dioxide and water gave more trouble than expected.

The WWTP and BLO systems started up May 16 and June 13, 1974, respectively. During and after the grant period, the WWTP provided 89 to 91 percent BOD_5 removal and effluent total suspended solids of about 80 mg/L. In the same period, the BLO system oxidized 98 percent of the sodium sulfide in the influent at an oxygen utilization efficiency of 75 to 80 percent.

Any data gathered after the grant period which refuted the period data are also reported. This report was submitted in fulfillment of Grant No. S 800740 by The Chesapeake Corporation of Virginia under the partial sponsorship of the U.S. Environmental Protection Agency. This report covers the period from June 19, 1972, to November 18, 1975, and work was completed as of November 18, 1975.

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LIST OF ABBREVIATIONS AND SYMBOLS

a.d.	—air dry
AO	—acid-oxygen bleaching
AODED	—acid-oxygen-chlorine dioxide-caustic-chlorine dioxide
ASP	—air separation plant
BLO	—black liquor oxidation
BOD_5	—biological oxygen demand after five days
CCV	—Chesapeake Corporation of Virginia
CED	—Cupriethylene diamine
CEDED	—chlorine-caustic-chlorine dioxide-caustic-chlorine dioxide
CIL	—Canadian Industries Limited
COD	—chemical oxygen demand
CSF	—Canadian Standard Freeness
D/C	—chlorine dioxide/chlorine
D/COD	—chlorine dioxide/chlorine-oxygen-chlorine dioxide
DO	—dissolved oxygen
FRP	—Fiberglass reinforced plastic
g/L	—grams per liter
gpm	—gallons per minute
hp	—horsepower
IMPCO	—improved Machinery Company
kg	—kilogram(s)
kg/cm^2	—kilgrams per square centimeter
km	—kilometer(s)
K-No.	—permanganate number
LEL	—lower explosive limit
LOX	—liquid oxygen
L/s	—liters per second
m	—meter(s)
mg/L	—milligrams per liter
min	—minute
OD/CED	—oxygen-chlorine dioxide/chlorine-caustic-chlorine dioxide
ODED	—oxygen-chlorine dioxide-caustic-chlorine dioxide
OUE	—oxygen utilization efficiency
pH	—hydrogen ion concentration
psig	—pounds per square inch (gauge)
SBH	—State Board of Health
SUPAIRCO	—Superior Air Products
SVP	—single-vessel process
SWCB	—Virginia State Water Control Board
TRP	—twin roll press
TRS	—total reduced sulfur
WWTP	—wastewater treatment plant
VIMS	—Virginia Institute of Marine Sciences

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ACKNOWLEDGMENT

We are particularly indebted to the late A. W. Plummer who, more than any other single person, was responsible for the application of oxygen technologies at Chesapeake. The other individuals involved in the projects described here are too numerous to list individually, but their involvement and contributions are hereby recognized and gratefully acknowledged.

SECTION 1

INTRODUCTION

The Chesapeake Corporation of Virginia began in 1920 to select obviously superior trees and leave them standing for natural reseeding and reforestation. The Company employed a professional forester and established its first seed orchard about 1936. In 1947, Chesapeake joined other companies in a widespread search conducted by North Carolina State University in the South and Southeast for naturally occurring superior pine trees. Cuttings from such superior trees grafted to growing stock in seed orchards produced increasing numbers of superior seeds. These were supplied to State nurseries and now provide enough superior seedlings to constitute half the reforested trees in Chesapeake's annual program.

Throughout these forty-odd years, hardwood standing in the timber procurement area has caused constant problems. For example, (1) hardwood composes about 50 percent of the standing timber, thus complicating normal harvesting of desirable softwood species, (2) short, stiff hardwood fibers were not suitable for Chesapeake's traditional and projected products, (3) limited utilization of hardwoods represented an inadequate use of a valuable, renewable natural resource, and (4) unless hardwoods could be harvested more extensively, reforestation in softwoods of average and superior species could not be expanded.

Seeking feasible approaches to increasing hardwood use, Chesapeake explored four major systems in depth between 1960 and 1971 and developed some marketable products with higher hardwood content. With the advent of partial or pure countercurrent washing in bleach systems, (1) it appeared possible for Chesapeake to undertake hardwood pulp bleaching with no increase in fresh water consumption. Thus in 1971, Chesapeake chose bleached hardwood market pulp as the most feasible approach to increasing the Company's use of hardwood. When making this decision Chesapeake considered the traditional uncertainty of the pulp market. The Company decided it must select a bleaching sequence that would consistently produce a high-quality market pulp at brightness levels of 88 or higher on the GE brightness scale.

Several conventional chlorination sequences, principally of 4 and 5 stages, were studied using run-of-the-mill hardwood as the wood source. The CEDED sequence (chlorine-caustic-chlorine dioxide-caustic-chlorine dioxide) was studied most intensively during in-house research in 1971.

Beginning in 1968, a factor in addition to increased hardwood use became dominant in Chesapeake's deliberations. The Virginia State Water Control Board (SWCB) informed (2) Chesapeake that it must install secondary wastewater treatment or its equivalent—or provide evidence that such treatment was unnecessary. Not having planned the hardwood expansion at that time, Chesapeake surveyed the Pamunkey River extensively to measure the influence of untreated wastewater discharge on the receiving water quality. A lengthly report (3) submitted to the SWCB concluded that little effect on water quality could be measured beneath and outside a floating and clearly definable mixing zone. Only a dissolved oxygen (DO) sag 1.4 to 3.2 km (1½ to 2 miles) upstream of the outfall remained in question.

To explain this DO sag, Chesapeake contracted with the Virginia Institute of Marine Science (VIMS) to conduct a dye-release study at the confluence of the Mattaponi, Pamunkey, and York Rivers. The VIMS report (4) following the study suggested that untreated wastewater discharged at the Chesapeake mill location could influence the sag. The report also stated that to maintain the minimum DO in the receiving waters at all times, Chesapeake would require secondary treatment only during the once-in-10-years, 14-consecutive-day low flow. Results of this VIMS study, financed by Chesapeake, became part of the public record and the Hyer-Fang mathematical model (5) for the entire York River basin.

Notwithstanding the findings of the 1968 river survey by Chesapeake and the dye-release study by VIMS in 1968—70, these results became academic when Chesapeake decided to expand production into bleached hardwood market pulp. The State Water Law was amended by the Virginia General Assembly in February 1970 to require secondary treatment or its equivalent of wastewater discharged if an industry expanded. Thereafter, Chesapeake started a search for the latest technology for wastewater treatment and included oxygen sequences in its bleaching research because of the potential for reducing bleach effluent. The Company also began engineering studies of in-process systems to reduce the pollution load that would have to be treated by an external wastewater treatment plant (WWTP).

Because of the numerous and interrelated tasks to be completed in the three systems of (a) in-process changes, (b) the expansion into bleached hardwood, and (c) the WWTP installation, Chesapeake completed the systems in that order. Sufficient in-plant reductions in waste load were made to offset the effluent load expected from the oxygen bleach system. This enabled Chesapeake to continue discharging waste water under Certificate No. 1260, issued by the SWCB in 1960, until startup of the WWTP targeted for 1 year after bleach plant startup. This sequence of events was approved (6) by the SWCB and permitted the investigation, pilot plant study, process design and installation of the latest and best technology, for Chesapeake, of high-purity-oxygen wastewater treatment.

Since Chesapeake perceived that total reduced sulfur (TRS) emissions and typical kraft mill odor reductions would be required in the near future, the overall systems using high-purity oxygen were broadened to include in-line weak BLO and possibly air enrichment at the lime kiln. An on-site oxygen plant was to supply the necessary low cost oxygen for all applications.

SECTION 2

CONCLUSIONS

The results obtained from the operation of the integrated oxygen processes and in-plant control steps allow us to draw the following conclusions:

1. The three-stage oxygen bleach system (D/COD) can produce a competitive quality bleached hardwood market pulp. However, without oxygen-stage (O-stage) effluent recycle, no reduction in BOD_5 and only limited reduction (36 percent) in effluent color was apparent.

2. With recycle and complete recovery of O-stage effluent, a 40-percent reduction in BOD_5 and a 65-percent reduction in color is possible.

3. With O-stage recycle, white liquor chlorides levels would be elevated from 6 to 17 g/L in a kraft pulp mill producing 20 to 25 percent bleached pulp.

4. Similar or larger reductions in BOD_5 and color would result from recycling effluents from an oxygen bleach system using oxygen in the first stage without potential corrosion problems from chloride build-up. (7,8)

5. The present bleach plant equipment, after necessary process and mechanical redesign, can successfully operate mechanically to produce bleached hardwood pulp. However, thickening pulp to 30 or greater percent consistency with a twin roll press results in high mechanical stresses. Equipment design must consider both these high stresses and the environment in which this equipment is to operate.

6. A twin roll press can operate successfully to thicken stock to 30 percent or greater ahead of the oxygen reactor at a pH of 4.5 and chlorides of 1000 mg/L.

7. The pressate from a twin roll press thickening hardwood pulp has high amounts of fiber and air, which makes recycling of this stream difficult.

8. Complete reuse of all condensates from the digesters and evaporators as brownstock shower water will result in a buildup of volatile BOD_5, primarily methanol, in the condensate and liquor systems. Sewering of these condensates, when necessary, will cause surges in BOD_5 to an external WWTP.

9. An on-site cryogenic oxygen plant, with liquid storage backup, can provide the low-cost oxygen necessary for continuous and economical operation of WWTP, BLO, and bleach plant systems. The total cost per ton of oxygen produced can be reduced incrementally by marketing excess liquid oxygen.

10. Cryogenic oxygen plants cannot efficiently produce at lower than design tonnage rates unless originally designed to do so.

11. Kraft mill wastes can be treated successfully with a high-purity-oxygen activated sludge system. When upset by short-term peak loads, return to stable operation is rapid. However, storage capacity is most desirable for spills of strong wastes, which can be fed back to the WWTP influent line at low controllable rates.

12. A high-purity-oxygen system will produce less excess biomass at higher solids concentrations than an air system. However, disposal of such biomass is very costly and difficult to achieve, especially when the ratio of primary to secondary solids wasted approaches 1:1.

13. A high-purity-oxygen, in-line BLO system will operate successfully to oxidize southern pine weak black liquors without foam or evaporator problems. The system is simple to operate, requires nominal space, and no special geometry for the oxygen reactor. Proper sizing of the pipeline downstream of the oxygen entry point is important to provide a high shear rate for the weak liquor oxygen mixture.

14. The use of oxygen for bleaching, WWTP operation, and BLO may be cost effective in separate or in combined installations at mill sites. All systems have now been commercially proved and should be viewed as alternatives to more conventional systems.

SECTION 3

RECOMMENDATIONS

Recommendations as a result of an evaluation of the operation of the integrated oxygen processes and in-plant control steps are:

1. If bleach plant effluent reduction is the primary goal, oxygen bleach systems with the O-stage first should be evaluated. Should Chesapeake decide to change bleach sequence, the Technical Department would recommend to Production and Company Management that the oxygen stage of the present system be repiped to be the first stage and that a caustic extraction stage be installed between the D/C and D-stage.

2. Oxygen bleach plants using a twin roll press should be designed with means to remove air from the pressate. Also the design should consider means to minimize pressate fiber loss. This is important to maximize recycle of pressate without causing washer operating problems.

3. A means of reducing the build-up of volatile BOD_5 in condensate and liquor systems, which is due to complete condensate reuse, should be considered in design of an in-plant effluent load control system.

4. The degree to which an onsite cryogenic oxygen plant needs to be turned down should be considered in initial design of the oxygen plant.

5. The transfer of biomass from the bioreactors to the secondary clarifier in the high-purity-oxygen waste treatment system should be accomplished with a minimum of turbulence and chance of air entrainment so as to minimize suspended solids in the final clarified effluent.

SECTION 4

MATERIALS AND METHODS

OVERVIEW

Many different materials and methods were utilized during the development and accomplishment of this project, since it was a large-scale commercial installation without significant full-scale precedent.

The materials of construction and the methods used to establish processes and equipment selection are reviewed briefly in this chapter under the following major headings: (1) In-process equipment changes, (2) Bleach plant equipment, (3) Air separation plant, (4) Wastewater treatment plant, and (5) Black liquor oxidation. Complete information on these topics is given elsewhere in this report.

Appendix A gives the materials of construction for all major items of equipment, including replacements for existing equipment, new units installed, and replacements where failures were experienced. Appendix C discusses problems encountered where failures occurred and the measures taken to alleviate them. Materials that were found to be unsuitable for the service are identified in Appendix C.

The methods used for selection of processes and equipment are discussed in Appendices A and C. The procedures for experimental analyses are thoroughly covered in Appendix B. The following chapter covers procedures for evaluation of the systems and equipment after installation.

MATERIALS USED

In-Process Equipment Changes

Three-Stage Vacuum Washers—
The three-stage vacuum washers were installed to wash the brown-stock from the digesters, replacing old screw press pulp washers. Typical pulp mill materials of construction were selected for this service: 304 stainless steel wire, steel vats, and steel drum support with stainless steel decking.

Steam Stripper—
The steam stripper was essentially a new item in the paper industry and the materials of construction were chosen from the limited experience available: 304 stainless steel was used extensively both as the primary material and as a lining for the steel shell of the main heat exchanger. The tubes of the reflux condenser were made of 316 stainless steel, while the tubes of the secondary condenser were made of XM33. The wet scrubber (using white liquor as the scrubbing medium) has a steel body and ceramic packing.

Surface Condensers—

The surface condensers receive and condense vapors from the last effect of the black liquor evaporator; the tube side is exposed to saline, screened river water, while the shell side contains the water vapor and black liquor condensate. Materials used were carbon steel for the condensate and black liquor contact, 316L for the saline river water contact.

Bleach Plant Equipment

Bleaching System Units—

The bleach plant was an entirely new type of installation and the materials of construction were chosen to withstand the special corrosive conditions in the oxygen bleaching system. For example, the oxygen reactor was built of mild steel but was lined with Carpenter 20 alloy. This material has given exceptional, trouble-free service. The first stage consisted of two mild steel tanks lined with fiber glass reinforced plastic (FRP). The third stage (D-tower), both updraft and downdraft, was fabricated of FRP alone. FRP was also used for the seal tanks throughout the bleach plant. The oxygen washer was built of 316 stainless steel with a polyester wire. The D/C washer was a titanium vat with a Kynar wire and the D-washer uses 317 stainless steel for the major components, with a polyester wire. The mixing towers were of carbon steel with an FRP lining.

Twin Roll Press—

The twin roll press was installed to thicken the bleached stock between the first and second stages of bleaching. This is a non-conventional item in a bleach plant and operates in a very corrosive environment not normally associated with this type of equipment. The original specifications called for a drilled 317 ELC stainless steel cover and a 317 ELC stainless steel shell. The actual installation utilized a 317 ELC stainless steel woven-wire cover in place of the drilled cover. The wire cover failed and was replaced with a drilled cover as originally specified. This new cover lasted about one year and then also failed. Changes were made to reduce the chloride level in the stock and the present roll cover and shell of 317 ELC stainless steel have been operated successfully since their installation.

Air Separation Plant

The air separation plant consists of standard manufactured products using materials of construction as determined by the manufacturers' experience. The oxygen distribution system uses an 853 carbon steel pipeline that was degreased and dried before start-up.

Wastewater Treatment Plant

The wastewater treatment plant consists primarily of large reinforced concrete structures such as sumps, clarifiers, the UNOX bioreactor, etc. The caustic strong waste storage tank was made of A-283 Grade C carbon steel and the kiln strong waste tank is also of carbon steel. The cooling tower is a commercially manufactured unit using conventional materials of construction for this service.

Black Liquor Oxidation System

The oxygen pipeline from the compressor in the oxygen plant to the spargers in the line was originally of low carbon steel, but was replaced by 316 stainless steel, which is the present service. The spargers are constructed of 304 L stainless steel. The pipeline from the spargers to the evaporators is of carbon steel and provides the required retention time and shear rate of reaction between the injected oxygen and the sulfide chemicals in the weak black liquor.

METHODS USED

In-Process Changes

The basic purpose of the in-process changes was to reduce the existing in-plant pollution load to a level that would permit operation of the new bleaching facilities without exceeding the applicable limitations on treated waste discharge. Methods were developed to achieve this purpose through the use of conventional engineering and scientific measurements and investigative programs. The methods used included surveys of the receiving stream and analyses of the various wastewater sources within the plant. Evaluation of the data led to decisions to replace certain process equipment units to achieve greater usage of existing water supplies through recycling and judicious conservation methods. Where this method was not sufficient, new pieces of equipment were specified and installed to provide additional recovery of pollutants from the existing waste streams. The net result of this approach was the reduction of existing in-plant wastewater to meet the overall objectives.

Bleaching System

The method used to select the bleaching system for installation in the mill consisted of extensive literature searches, in house experimentation, processing of samples by outside organizations, and negotiations with prospective suppliers of the bleaching equipment. Many systems were investigated, and when the oxygen system became prominent in the research, Chesapeake personnel visited pilot plant operations in Sweden and South Africa for first-hand observations. The work of many organizations was reviewed and the ultimate selection was based upon compliance with the wastewater treatment load generated, the cost of installation and operation, and the effectiveness of the process to produce marketable products.

This method of operation was conceived and directed by Chesapeake personnel and proved to be effective in achieving the objectives.

Air Separation Plant

The method used to select the air separation plant consisted of examining the various types available from standard commercial manufacturers and requesting bid tenders from those who appeared to have the most suitable systems. The choice was narrowed down to two types of systems and the ultimate selection of a cryogenic system was based upon the higher oxygen purity produced by this equipment, as well as the lower capital and operating costs for the production rate.

Wastewater Treatment Plant

The methods used to develop the wastewater treatment plant began by retaining highly qualified experts in the field and following their advice from theoretical concepts through pilot plant operations to the full-scale installation. Laboratory trials were run on a small scale to obtain data for design of the full-scale plant. Specialists in various phases of design and equipment were contacted and their experience was ultilized in equipment selection. An engineering firm was then hired to develop capital cost estimates for alternative systems.

The ultimate selection of process and equipment was based upon the system's ability to meet the removal efficiencies, operate effectively and reliabily over long periods of time, and to be cost effective as compared with other technical approaches. The results of these studies and comaprisons were then utilized as the basis for design of the system, selection of equipment, and construction of the plant.

Black Liquor Oxidation System

The black liquor oxidation process was selected based primarily upon experimental work done at Chesapeake. Extensive bench scale tests, as well as a mill-scale trial of one month's duration, were performed. Data gathered from these studies was used to develop capital and operating costs, which were incorporated into the overall waste treatment system cost and evaluated for economic effectiveness versus competitive methods. The present black liquor oxidation system was then design and installed along with the other waste treatment facilities.

SECTION 5

PROCEDURES

Procedures used and steps taken to select, install, and evaluate processes using high-purity-oxygen for bleaching hardwood pulp, wastewater treatment, and BLO involved numerous scientific and engineering disciplines and were conducted both in-house and externally at wide-spread locations. This overall work differed considerably in many respects from traditional experimental procedures used in a limited (or smaller scope) scientific or engineering project, although such procedures were involved in each phase. The approaches followed are described briefly below and are amplified in Appendices B and C. Although not included in the grant period, certain pre- and post-period information is included to demonstrate the rationale for decisions made.

IN-PROCESS CHANGES

The Virginia SWCB was organized in 1946. In its first decade, emphasis was placed on statewide receiving water surveys to establish minimum ciriteria and quality standards. Beginning in 1957, Chesapeake made summer surveys of the Pamunkey, Mattaponi, and York Rivers over a 40.2-km (25-mile) stretch, measuring BOD_5, pH, salinity, temperature, and DO. These data were collected to provide background information on receiving water quality over a long period of time. The surveys were made annually except during the first 2 years after the WWTP started up in 1974.

Chesapeake planned an expansion in mill production to take place in 1964-65. SWCB approval was granted, and discharged Certificate No. 1260 was issued to Chesapeake in June 1960, permitting a BOD_5 daily discharge of 19,068 kg/day (42,000 lb/day). Thereafter, Chesapeake measured several parameters of process wastewater streams and reported the results quarterly to the SWCB. These were mandated reports, but the data assembled proved invaluable for decision on effective in-process waste stream reductions. These reductions would allow Chesapeake to operate a bleach plant for about a year before startup of a WWTP—without violating the 19,068 kg BOD_5/day (42,000 lb/day) requirement of Certificate No. 1260. The principal purpose of the delay in starting up the WWTP was to allow time to evaluate high-purity oxygen treatment of wastewaters. The SWCB approved (6) Chesapeake's proposal for the postponement so long as the terms of Certificate No. 1260 were not contravened.

The in-process changes made were (a) replacement of screw press washers and cold stock screening with hot stock screening, refining and 3-stage vacuum washers, (b) installation of a steam stripper to strip certain evaporator and blow heat condensates and turpentine decanter underflow, (c) replacement of direct contact jet condensers with surface condensers on all three sets of multiple-effect evaporators, and (d) installation of improved entrainment separators in the evaporators. These changes are discussed in Section 6.

To be certain that enough fresh water was available for adding a hardwood bleach system, Chesapeake intensively traced pipelines, measured flows and revised the pulp mill water system extensively. The amount of water required for the bleach system was reduced by considering only an overall countercurrent system from the pulp dryer back to the two bleach plant washers. Also, within the bleach plant itself filtrates were recycled where feasible to reduce water requirements.

BLEACHING SYSTEM

Sequence Selection

In-house bleaching research concentrated on numerous conventional chlorination sequences. Most effort finally focused on the CEDED sequence. Laboratory procedures used were standard or typical of traditional laboratory bleaching sequence studies. These are detailed in Appendix B.

Being mindful of the need to keep bleach effluent loads low, Chesapeake took note of two so-called pilot oxygen bleaching sequences at Enstra, South Africa, and Husuum, Sweden. Chesapeake technical and operating people visited both locations. Chesapeake collected and evaluated the properties of pine pulp before and after the AO stage of Enstra and the O_2 stage at Husuum. Chesapeake also sent samples of unbleached hardwood pulp to L'Air Liquide, Paris, France and, through Chemetics International, Vancouver, B.C., to Canadian Industries Limited (CIL), McMasterville, Quebec, for laboratory bleaching trials of various oxygen sequences. L'Air Liquide performed AO bleaching of the samples, and Chesapeake completed numerous sequences on the AO bleached pulp with emphasis on AODED. CIL investigated COD, D/COD, OD/CED, D/C ODED and CEDED. D/COD was evaluated for both hardwood and softwood. Chesapeake studied the D/COD sequence in-house by use of a pilot batch digester rebuilt to serve as an O_2-reactor. Appendices B and C expand on how Chesapeake took these steps.

Process Design, Engineering and Installation

Process design of the sequence, D/COD, was based on research done at Chesapeake and CIL. Chesapeake then negotiated a turn-key contract with Chemetics International as the prime contractor to engineer and install the bleach plant. Several other major vendors were involved such as Improved Machinery Company (IMPCO), Nashua, N.H. and Hooker Chemical Company, Niagara Falls, N.Y. who supplied the single-vessel process (SVP) chlorine dioxide generator. Natkin Construction Company, Richmond, Virginia erected the bleach plant under contract to Chemetics and the SVP system under contract to Tidewater Construction, Norfolk, Virginia.

Startup Preparations—
Chemetics provided an operating training manual and conducted extensive training classes for both production and technical employees. Chesapeake's Research Department prepared a testing manual for bleach plant operators and for appropriate Technical Department personnel.

Bleach Plant Evaluation

Chesapeake's initial and most intense studies after bleach plant startup were in this order, (a) how to solve extensive mechanical problems, both major and minor, (b) what quality bleached hardwood pulp could be produced with the D/COD sequence, (c) what effluent loads flowed from the plant, by stage and combined and (d) what effect would recycling the O-stage filtrate have on the chemical recovery system.

The procedures employed in item "a" were legion, or, as mechanical difficulties occurred judgments were made on the best solutions. In numerous instances several changes were necessary before the problem was solved. Appendix C gives specific details of both problems and solutions.

Quality of the bleached pulp produced by the three stages was checked by measuring a number of product parameters. Major time was spent studying brightness, or whitness, of the pulp; the extent to which the process removed dirt in the form of bark specks, transient particles and shives, or relatively uncooked splinters of wood, and what changes occurred in pulp viscosity. The procedures used in measuring pulp quality are described in Appendix B.

The combined effluent from the bleach plant was sampled and measured for BOD_5 and color on a regular basis from February 1973 to April 1974. Chesapeake reported (9) the combined effluent data for the bleach plant for 1973 to EPA and WAPORA, INC., in March 1974. Stage-by stage effluent was measured for BOD_5, color, sodium, chlorides, and flow.

After the grant period (November 18, 1975), it appeared that effluent loadings from the bleach plant to the WWTP were higher than those measured and reported during the grant period. Therefore, an additional effluent material balance was made by taking daily composite samples of each of the seal tank overflows and the combined effluent for one week. BOD_5, color, sodium, chlorides, and flow were measured on composite samples. Data collected are summarized in Section 6 and provided in detail in Appendix C. Using a computer technique for solving simultaneous equations, flows for each stage were estimated. The bleach plant effluent meter was also checked using the lithium salt tracer method. Details of the procedures used are in Appendix B.

Recycling O-Stage Effluent

The effect of recycling O-stage effluent was determined by using the chloride levels for the O-stage determined by the 1976 effluent by stage study. These data were used in conjunction with a pulp mill sodium and chloride balance to determine white liquor chloride levels if the O-stage effluent was recycled.

AIR SEPARATION PLANT (ASP)

Selecting Type of Plant

Two basic types of ASP's were considered, namely, (a) pressure swing adsorption (i.e., molecular sieve) and (b) cryogenic. Proposals were obtained on both types starting at a plant capacity of 36.3 metric tons/day (40 tons/day) but eventually for 90.7 metric tons/day (100 tons/day) as additional uses for oxygen were considered and adopted. Several factors influenced the choice of a cryogenic ASP. These are discussed in Appendix C.

Plant Design, Engineering and Installation

Plant design and engineering for the ASP was provided by Superior Air Products, (SUPAIRCO), Sayreville, N.J. Installation of the ASP was supervised by SUPAIRCO and Rust Engineering. Actual erection of the ASP fell within the turnkey contract with Rust Engineering for the WWTP reported below.

Startup Preparation

A representative of SUPAIRCO spent several weeks at Chesapeake before the ASP startup. He trained operating, maintenance and technical employees to carry out their normal functions on a system new to Chesapeake. Individual equipment in the ASP was test run sequentially before overall startup. The SUPAIRCO representative, along with SUPAIRCO's chief engineer, assisted and supervised startup of the ASP and remained at Chesapeake until the plant was operating stably.

ASP Evaluation

The ASP was evaluated after startup to establish whether or not the plant would produce the designed 90.7 metric tons/day (100 tons/day) of oxygen 81.6 metric tons/day (90 tons/day) gas at 172 kPa (25 psig) and 9.1 metric tons/day (10 tons/day) liquid at a purity of 99.5 percent or higher with a methane content of 0.25 percent or less. Results of this evaluation are included in Section 6 and Appendix C.

WASTE WATER TREATMENT PLANT (WWTP)

Process Selection

Chesapeake explored several options for wastewater treatment beginning early in 1967. For assistance in evaluating and comparing various treatment systems Chesapeake later (October 1971) retained Hydroscience, Inc., Westwood, N.J., widely recognized specialists in the subject area from theoretical concepts to full-scale domestic and industrial systems.

Chesapeake purchased and set up for laboratory trials a small-scale air activated sludge system. This was operated to establish treatability data and oxygen uptake rates. The system was too small to provide quantitatively reliable data or to produce enough sludge to measure generation rates, settleability, and dewatering characteristics. Air activated sludge treatment data compared with that from other treatment systems, as indicated below, was largely supplied by Hydroscience, Inc. from their considerable experience.

Chesapeake contacted the Linde Division, Union Carbide Corporation, immediately after initial publicity appeared on the Batavia, New York UNOX system (early in 1970). Chesapeake sent technical and engineering personnel to Batavia to observe plant operation and control. Several trips were made to Tonawanda, N.Y. to see equipment and negotiate with the Linde Division the purchase of a mobile trailer-mounted UNOX pilot plant of 1.9 L/s (30 gpm) capacity. The contract was signed in March 1971 and the pilot plant operated by Chesapeake from September 1971 to April 1972 on four mill waste streams (not including bleach effluent since none was available).

Process Design, Engineering and Installation

Results from pilot plant operation in several modes were used by Chesapeake, Union Carbide, and Hydroscience, Inc. to provide process design and estimated capital and operating costs for a full-scale UNOX waste water treatment system. Similarly, Hydroscience, Inc. provided process design and estimated capital and operating cost data for an aerated lagoon and air activated sludge system for comparison with the UNOX system. All process design data were used by The Rust Engineering Company, Birmingham, Alabama (Rust) to provide capital cost estimates for all three systems.

Chesapeake computed the net present value after tax, cash flow cost for the wastewater treatment systems alone and included BLO using air or oxygen in a separate computation. Process system selection was based on these computations (with air and oxygen BLO) plus certain intangible factors discussed in Appendix C. Approval of process design, with terms and conditions, was received from the State Water Control Board (SWCB) in November 1972.

Engineering and installation were handled by Rust on a turnkey contract but in close consultation with Chesapeake on engineering and P&I diagrams. Construction started on March 17, 1973, two weeks behind schedule because of inclement weather.

Startup Preparations

Chesapeake applied to and received from the U.S. Army Corps of Engineers, and appropriate state and local regulatory agencies, permits to install a diffuser in the Pamunkey River for the effluent discharge from the WWTP.

Similarly, requests for approval of two alternate landfill sites were made to and received from the State Board of Health (SBH) and SWCB for disposal of primary and secondary sludge from the WWTP.

Chesapeake personnel attended a one week operator training course. After substantial further preparation a supervisor who was to have responsible charge of the WWTP operation took and passed the operator certification examination.

Chesapeake conducted training classes for operators, maintenance people, and power plant supervisory people. Also, numerous meetings were held with maintenance people to select spare parts. Orientation meetings were held with operating department superintendents and other operating supervisors to describe the WWTP startup and the operation of the effluent collection and spill prevention system.

Numerous piping, mechanical, and pump delivery problems were experienced and corrected during preparation for startup. Some caused major delays but with all the problems, large and small, the primary clarifier system and sludge vacuum filters were started up April 23, 1974. The bioreactor and secondary system was seeded May 16—17, 1974.

A joint SWCB-EPA NPDES permit, issued to Chesapeake in June 1974, included January 1, 1975 as the required compliance date. Because of the long delayed delivery of sump pumps, especially for the bleach plant, Chesapeake requested and received a delay in the compliance date to August 1, 1975 with requirements to report periodically on construction program progress.

Well before WWTP startup Chesapeake developed operating logs and planned sampling and testing procedures for WWTP evaluation and collection of data for mandatory reporting monthly to the SWCB.

WWTP Evaluation

Evaluation of the WWTP was delayed frequently by mechanical problems and by liquor and turpentine spills to the WWTP collection systems. Throughout the period June 1974 to August 1975 Chesapeake monitored input, output, and internal parameters of the WWTP and reported operational results to the SWCB, including performance, problems, and steps taken to correct the problems. Not until January 1975 was bleach plant effluent available for treatment and for sometime thereafter the WWTP was monitored frequently without bleach effluent because of bleach plant sump pump problems.

Because of sump pump plugging and long delivery of replacement pumps, complete and continuous data on the WWTP could not be assembled until late 1976. In fact, the most satisfactory and complete set of information was collected in mid-1977. The latter data are reported in Section 6; more detail and discussion is provided in internal reports and correspondence discussed in Appendix C.

BLACK LIQUOR OXIDATION (BLO)

Process Selection

Exploratory laboratory work was done at atmospheric pressure to oxidize both weak and strong black liquor with oxygen. Based on the promising results of this work a mill-scale trial was run for one month on one set of quintuple-effect evaporators. Gaseous oxygen injected into the weak black liquor before entering the first effect was steam-vaporized-liquid-oxygen stored onsite. Following this trial a smaller pilot BLO system was operated in the evaporator room to establish system pressure and reactor shear rate to optimize oxygen utilization efficiency (OUE).

No laboratory or pilot plant work was done on air oxidation of strong black liquor. However, Rust Engineering's experience in conjunction with a literature review was used to provide capital and operating cost estimate inputs to the comparative net present value computations referred to above.

Process Design, Engineering and Installation

Process design of the weak BLO system was based on results of the mill-scale and pilot trials. Rust engineered and installed the full-scale BLO equipment and control system.

Startup Preparations

A number of changes were made in the BLO oxygen liquor system as each part was test run. These changes are reported in Section 6 and Appendix C.

BLO Evaluation

After the changes mentioned above had been completed the BLO system was operated throughout a test period. No attempt was made to control operating parameters at constant or fixed levels. The system was allowed to float on stream and to respond to normal fluctuations. During this test period crucial system parameters were measured. These included (a) weak black liquor flow, pressures, and temperatures, (b) oxygen flow and (c) inlet and outlet sulfide concentrations in the weak black liquor. Oxidation reversion through the evaporators was also measured. Procedures used are described in Appendix B.

During this test period normal variations in the measured parameters spanned a range large enough to permit several parametric correlations. From the correlations the relative influence of the parameters on oxygen utilization efficiency (OUE) was established.

Post-Grant Period

After November 18, 1975, the termination date of the grant period, a BLO pilot unit was installed in parallel with the full-scale system. The purpose was to explore, and improve, what appeared to be a lower-than-expected OUE in the full scale system. The parameters measured in the full scale system were also measured in the pilot plant. The shear rate range explored in the early BLO pilot work was extended to 3 x 10^6 min $^{-1}$ in the post-grant period pilot trial.

SECTION 6

RESULTS AND DISCUSSION

This section of the report presents summaries of the performance and major operating problems with the following systems:

1. In-process
2. Bleach plant
3. Air separation plant
4. Wastewater treatment plant
5. Black liquor oxidation system

Operating results and data reported were gathered when the kraft pulp mill was averaging about 952 metric tons/day rate (1050 tons/day) and the bleach plant about 213 metric air dry tons/day (235 tons/day). The pulp mill production was nearly equally split between a Kamyr digester line producing 100 percent pine and a batch digester line alternating between pine and bleachable grade hardwood. Total mill production of unbleached linerboard, unbleached paper, and bleached hardwood market pulp was some 1066 metric tons/day (1175 tons/day). The additional 113 metric tons/day (125 tons/day) product above the pulp mill production level was supplied by use of secondary fiber.

Detailed results and discussion, as well as description of equipment is included in Appendices A and C. Where appropriate, pre- and post-grant period data and discussion is also presented.

IN-PROCESS CHANGES

In-process changes were made to reduce BOD_5 and flow to an external WWTP and to ensure compliance with existing permitted discharge limits, after bleach plant startup, and before the WWTP startup. Also, process changes were made in the pulp mill water system to supply sufficient water of proper temperature for the new bleach plant.

Description of In-Process Changes

A 3-stage vacuum washer was installed to replace three older, less efficient screw press washer systems. Also evaporator surface condensers and centrifugal catchalls were installed on three sets of evaporators. The surface condensers replaced direct contact condensers and the centrifugal catchalls removed sufficient additional black liquor entrainment to allow reuse of surface condensate as brownstock washer shower water. Lastly, a steam stripper was installed to remove volatile BOD_5 and odors from the turpentine decanter underflow. This system was designed ultimately to strip the condensates from the batch digester blow steam and the noncondensible gas condensers on the evaporators. Stripped condensate was to be used as brownstock washer shower sater. The volume and BOD_5 levels in the condensate streams to be stripped are summarized in Table 1. A flowsheet of the stripper column is shown in Figure 1.

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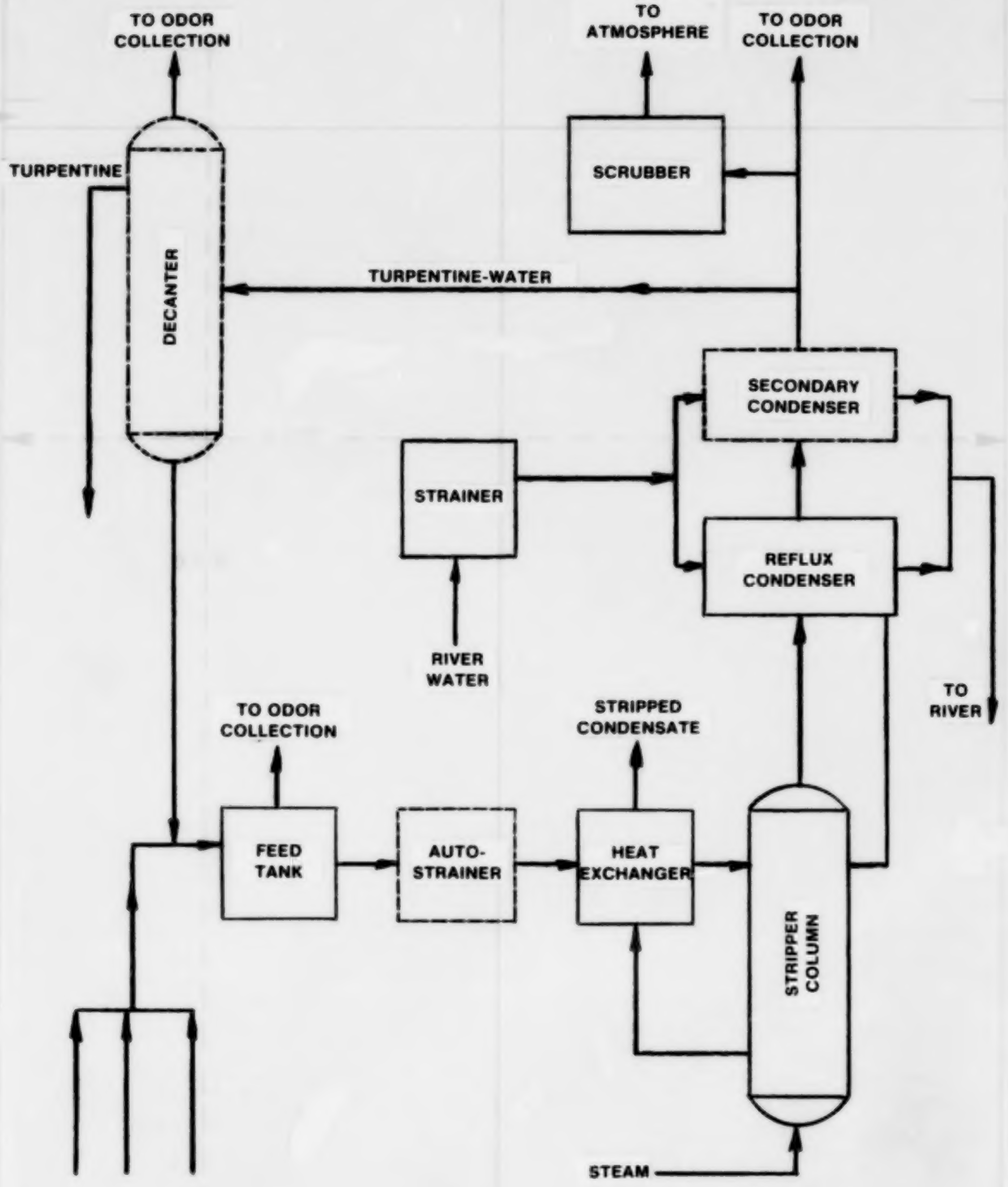

Figure 1. Steam stripper flowsheet.

TABLE 1. VOLUME AND BOD_5 CONTENT OF CONDENSATE STREAMS

Source	Volume L/metric ton (gal/ton)	Flow L/s (gpm)	BOD_5 Content kg/metric ton (lb/ton)	kg/day (lb/day)
Turpentine decanter underflow*	417 (100)	4.6 (73)	3.5-4 (7—8)	3,583 (7,900)
Evaporator noncondensible gas condensers*	41.7 (10)	0.46 (7)		
Batch digester blow steam #	1252 (300)	6.87 (109)	7.5 (15)	3,583 (7,166)
Total		11.93 (189)		7,166 (15,800)

* Based upon total pulp production of 952 metric tons/day (1052 tons/day).

# Based upon digester batch pulp production of 476 metric tons/day (525 tons/day); volumes shown are per batch-ton.

Operating Problems and Results

The brownstock vacuum washer and the evaporator surface condensers-centrifugal catchall combinations were conventional equipment installations. As such these systems presented only normal startup problems. The steam stripper, however, was a new piece of process equipment in the North American pulp and paper industry. The only known installations at the time were those in Europe and in the new Mead mill at Escanaba, Michigan.

Initially, the stripper system was to treat turpentine decanter underflow and the condensates from the evaporator noncondensible gases. The stripper gases were to be vented directly to the atmosphere after the reflux condenser. Therefore, the equipment shown in dashed lines in Figure 1 was not initially installed.

The column, in this operational mode, effectively removed BOD_5 (60 to 95-percent) and odors (80 to 99 percent) from the decanter underflow and evaporator condensates with a steam usage of 5 to 10 kg per 100 kg feed. (10) However, the odors emitted from the column were of such intensity as to require a scrubber using white liquor to be installed on the vent gases. This scrubber reduced the odors sufficiently to allow stripper operation.

A second significant operating problem occurred in the stripper feed heat exchanger. This unit was plagued with gasket failures and tended to plug. The rubber gaskets in the original equipment were replaced several times. After continued gasket failures, a spiral heat exchanger was installed to replace the original plate and coil heat exchanger. This solved both the plugging and the gasket failure problems. Other stripper operational problems were minor.

After the completion of the grant period a batch digester blow-steam condensing system and a non-condensible odorous gas collection and incineration system was installed in the pulp mill. To alllow steam stripping of the blow steam condensate and incineration of the stripper off-gases, several changes were made to the stripper column. These included:

1. An automatic strainer to remove small chips or fibers from the feed stream.
2. A secondary condenser to partially remove water vapor and turpentine prior to incineration.
3. A turpentine decanter to collect any turpentine that was stripped.

The equipment added during these changes were indicated by dashed lines in Figure 1. Collection blow steam condensate resulted in a final combined flow to the column of 12 L/s (190 gpm).

An estimate of the overall effect of the in-plant changes undertaken was to reduce flow to an external treatment system by 442 to 631 L/s (7 to 10,000 gpm) and BOD_5 by 4500 kg/day (10,000 lb/day). It also provided about 12 L/s (190 gpm) of condensate suitable for reuse as brownstock shower water. This allowed complete reuse of all digester and evaporator condensates. Other changes in the pulp mill water system provided the necessary 1500 gpm of water to supply the estimated requirement of 29,200 to 31,300 L/a.d. metric ton (7000 to 7500 gal/a.d. tons) for the 272 a.d. metric ton (300 a.d. tons) bleached plant operation.

All of the in-plant systems discussed above are still operating effectively today. The complete recycle of digester and evaporator condensate, however, has caused a buildup of volatile BOD_5, principally methanol, in our condensate and liquor system. Sewering of these condensates, when liquor carryover results, causes surges in BOD_5 to the external WWTP. Changes in the stripper column system to purify methanol to the point that it can be economically burned are being considered. This will result in more efficient methanol and BOD_5 removal and a lowering of methanol concentrations in all condensate streams.

BLEACH PLANT

The oxygen bleach plant started up in December 1972 and was the first oxygen system to begin operation. The oxygen bleach plant was selected since it was projected to produce a competitive, if not superior pulp, at lower overall costs and with less organic and inorganic pollutant discharges. The D/COD sequence selected to produce 88+ brightness pulp was and still is unique in the pulp and paper industry. The sequence of events, leading up to the selection of this sequence was detailed in Section 5. Studies undertaken to select the sequence are detailed in Appendix C.

Bleach Plant Description

The bleach plant was supplied by Candian Industries Limited (CIL) on a turnkey basis. There have been vital process changes since the original plant was started up, however, the basic sequence of D/COD is unchanged. Figure 2 illustrates the present bleach plant flow.

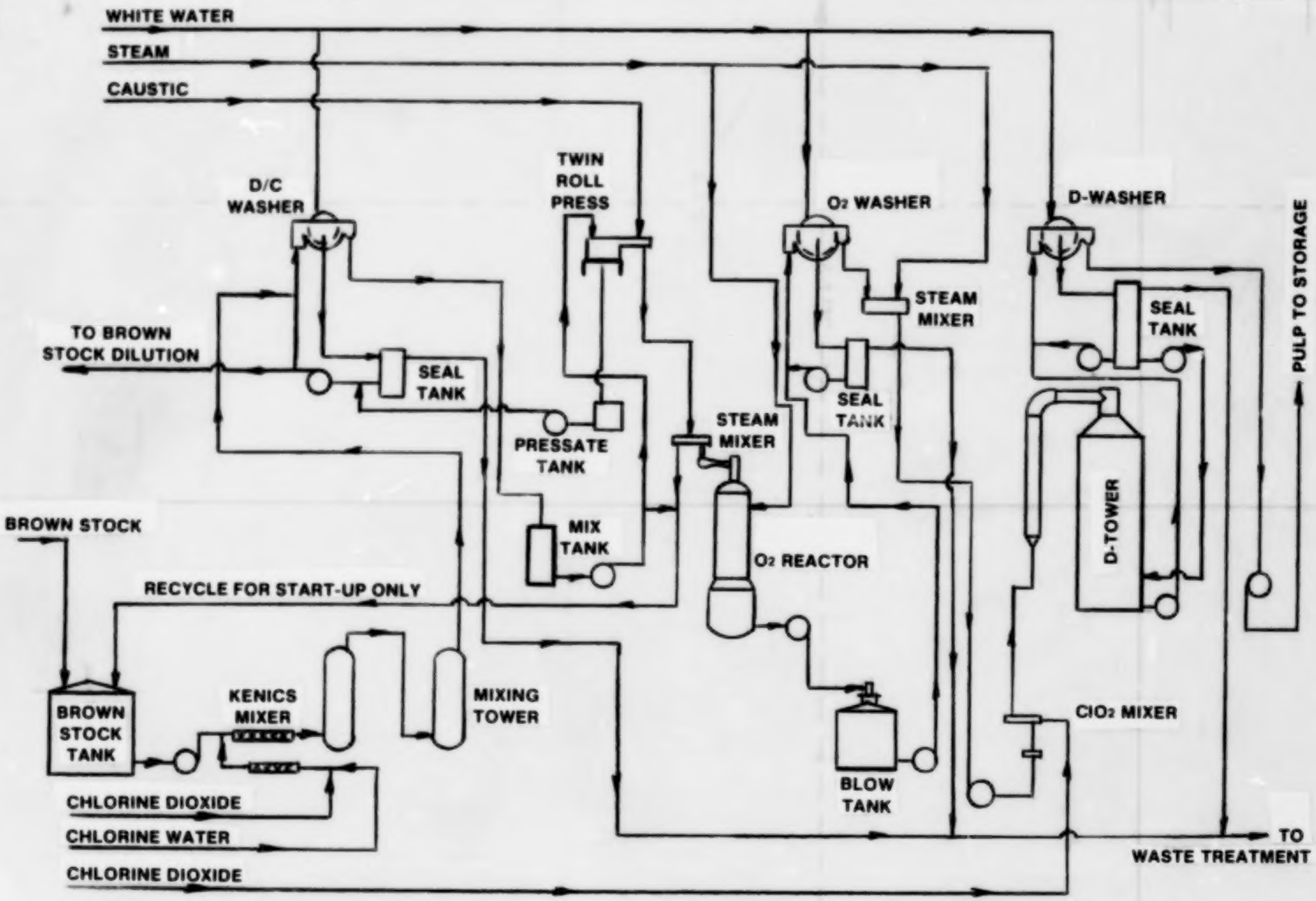

Figure 2. Bleach plant sequence.

The D/C stage is where an aqueous mixture of chlorine dioxide and chlorine is mixed intimately with unbleached pulp (brownstock). The chlorine comes from two sources. Chlorine water, a byproduct of manufacturing chlorine dioxide, is mixed with chlorine gas by means of an eductor. The strong chlorine solution is then mixed with the chlorine dioxide solution. The presence of chlorine dioxide moderates and inhibits the normally drastic reactions of chlorine alone at the high, first-stage temperature. The pulp at 3 to 4 percent consistency (percent dry fiber) is mixed with the chemicals in a Kenics in-line static mixer. Pulp and chemicals pass through two IMPCO Chloromix towers in series which provide retention time of 5 min each. The temperature in this stage is about 52°C (125°F).

The chlorinated pulp leaving the towers is diluted to 1 percent consistency with recycled washer filtrate, then washed on a vacuum washer with dryer machine white water. Pulp leaves the washer at a consistency of 13 percent.

The washer is constructed of corrosion resistant materials, the vat being titanium lined and the drum made mostly of FRP (Fiberglass reinforced plastic). In the original bleach plant, with no D/C stage washer, the chlorinated pulp was fed directly to an IMPCO Vari-Nip twin-roll press. The press rolls were continuously stresed in a highly corrosive environment. Due to these conditions stress-corrosion cracking occurred in the press rolls. The object of the new washer is to remove chlorides and increase pH to protect the twin-roll press.

From the D/C washer the pulp passes to the twin-roll press at 3 to 5 percent consistency. The press discharges pulp at 32 to 35 percent consistency to a breaker conveyor where a solution of 7 percent sodium hydroxide is added. The stock drops into a single shaft mixer and from there to a Sunds screw feeder. The pulp compacts in the feeder and serves as the pressure seal before entering the pressurized oxygen reactor. At the top of the reactor, it is fluffed by a Sunds-Impco device to increase the surface area for oxygen transfer.

The reactor is 3.05 m (10 ft) in diameter and 12.2 m (40 ft) high. The upper section of the reactor contains the bed of fluffed pulp. Pulp moves from the upper section to the lower where it is diluted and mixed by a rotating rake before blowing (discharge). Oxygen enters the reactor at the top of the reactor to maintain a pressure 414 kPa (60 psig). Since under certain conditions there can be a high fire risk in the oxygen reactor, various safety features were part of the initial design. If the exothermic reaction, or any other cause increases temperature excesively, pulp in the reactor is automatically showered with water. An automatic vent relieves excessive pressure. If this does not drop the pressure quickly, or fails to operate, a spring-loaded safety valve and two rupture discs are provided to relieve the excessive pressure. After a retention time of 30 to 45 minutes in the oxygen reactor, the pulp is discharged through a blow valve to a blow tank at 4 percent consistency.

The oxygen bleached pulp passes across a vacuum drum washer constructed of 316 stainless steel. The oxygen washer filtrate is reused for dilution purposes in the blow tank and oxygen reactor. The washed stock at approximately 16 percent consistency goes to a steam mixer and then to a mixer where chlorine dioxide is added. The mixed thick stock and chemicals are pumped to the up-draft tube of the chlorine dioxide tower [5.5 m (18 ft) diameter, 27.5 m (90 ft tall)]. Total retention time in the tower is 4 to 5 hours. The pulp is washed again and thickened on a vacuum drum filter constructed of 317 ELC stainless and pumped to a 181-metric-ton (200-ton), high-density tile storage tower.

Operating Problems

Major operating problems occurred on startup of the bleached plant. Most problems centered around the two non-conventional pieces of equipment, namely the twin roll press and the oxygen reactor. Problems with other equipment are typical of those experienced in conventional bleach sequences.

Twin Roll Press—

The most operating problems occurred with the twin roll press. Three roll failures occurred due to a combination of stress-fatigue and stress-corrosion cracking. The press rolls were initially in a highly stressed and very corrosive environment (pH of 1.6 to 2 and 1500 to 2000 mg/L chlorides). The outer shell of the original press rolls failed in 10 days. The original press rolls were 317 stainless steel, the outer shell being a "crimp back" wire, 3.2 mm diameter (⅛-in.). The crimp in the wire created numerous stress centers resulting in rapid and multiple failures of the roll covers. This type cover was replaced with drilled 317 stainless plate covers which failed after one year of service. At this time it was also discovered that the sub-structure of the press rolls was also cracking. To resolve this problem a D/C stage washer was ordered. The new washer has reduced chloride levels at the press by about 50 percent and has allowed economical press pH control at a 4.5 level. Under these conditions the present rolls and roll covers have been operating successfully for the last four years. The D/C washer is not highly stressed and is built of FRP and titanium where a corrosive environment exists. It has operated without significant corrosion problems to date.

The new D/C washer was expected to be more efficient at removing chlorides. However, good operation of the washer was, and still is, hindered by the high fiber and air content of the pressate from the twin roll press. High fiber and air content in the pressate has been a consistent and unexpected problem since bleach plant startup. Mill trials on the D/C washer have demonstrated that it is best to recover the fiber from the pressate by recycling it to the washer vat rather than using it as washer shower water. (11) New systems using a twin roll press should be of sufficient size to allow lower vat pressures so as to minimize fiber losses. Also the design should incorporate larger seal tanks or other proven methods to remove air from the pressate.

Oxygen Reactor—

The major operating problem with the oxygen stage has been with the drive on the oxygen reactor discharge assembly. Spalling of parts of the original hydraulic motor caused substantial production loss as did gear breakage on a new mechanical drive system which was installed to replace the hydraulic drive. Also an upset in operating conditions caused pulp to fill the bottom of the reactor and lift the original reactor discharge table. As the pulp was removed from the bottom of the reactor, the discharge table dropped unevenly and caused damage to the table and drive.

In the fall of 1976 the initial discharge table was replaced with the rake type presently used. This new type rake offered the options of operating at longer (almost double) retention times in the reactor or increasing production rate at the original retention time of 30 min at 272 a.d. metric tons/day (300 a.d. tons/day). Furthermore, maintenance of the rake type discharge has been negligible.

Initial problems were also experienced with operation of the oxygen washer and seal tank. Washer mat formation and pulp discharge consistencies were poor. Also considerable foam was generated by the washer seal tank. The basic cause of other these problems was due to solids build-up in the O_2 seal tank due to insufficient shower water being applied to the O-washer. In the particular case of the O-stage washer more water than normal needs to be added as shower water to operate the system at desired seal tank solids levels. This is because pulp enters the oxygen reactor at 28 to 30 percent consistency and leaves the washer at about 16 percent. This consistency change results in an additional water requirement. Without the addition of sufficient shower water to operate at a positive dilution factor, a buildup of dissolved solids in the seal tank had to occur. When this happened, seal tank foaming and poor washer operation resulted. The problem was improved dramatically by increasing shower water on the washer to obtain a positive dilution factor.

Pulp Quality Measurements

All pulp produced by the oxygen bleach plant was to be sold on the open market. Therefore, good pulp quality was a necessity. The most significant pulp quality problems were centered in two areas. These were first, achieving acceptable viscosity and strength levels, and second, achieving acceptable shive levels in the final bleached pulp.

Improving Pulp Viscosity and Strength—

Problems initially existed with obtaining high pulp viscosities and associated pulp strength levels at 300 Canadian Standard Freeness (CSF). Investigations revealed that the problems were two-fold: (1) the unbleached pulp fed to the bleach plant had a lower viscosity than the initial, experimental pulp and (2) there was about a 60 percent loss in viscosity across the bleach plant. The high viscosity drop in the bleach plant was occurring primarily in the high temperature D/C stage and the O-stage.

After a long series of laboratory and mill tests involving unbleached as well as bleached stock preparation three major conclusions were drawn:

1. Unbleached viscosity levels could be increased by raising unbleached K-No. (permanganate No.) levels and by minimizing the pine content in hardwood digester cooks.
2. The viscosity drop in the D/C stage could be minimized by controlling the brownstock pH into this stage and by the addition of blended chlorine and chlorine dioxide instead of sequential addition of chlorine dioxide and then chlorine.
3. The addition of magnesium sulfate ($MgSO_4.7H_2O$) ahead of the O-stage significantly reduced the viscosity drop in that stage.

As far as possible, operating procedures and process changes were made in the bleach plant to implement the conclusions drawn in the laboratory and mill trials. As a result, brownstock viscosities improved and viscosity drop through the bleach plant was reduced from 60 to 65 percent to 30 to 40 percent. Currently we are bleaching a 16-K-No. pulp with blended chemicals at the D/C stage and using D/C filtrate to provide some of the brownstock dilution needs. This drops, but does not always control pH in the best range (3.0 to 3.5). We are also adding magnesium sulfate ahead of the oxygen stage at a 3.7 kg/metric ton (7.5 lb/ton) level based on the weight of $MgSO_4.7H_2O$.

Viscosity levels after these process changes have been maintained at competitive market pulp levels. Also our bleached pulp physical strength tests (burst, breaking length, and tear) when measured at the same viscosity and pine content have compared favorably with competitive pulps. (13)

Shive and Dirt Level Measurement—

Shive levels in the final bleached pulp became a problem early in 1975 when the bleached hardwood pulp market became very tight. In examining the shives contained in the bleached pulp it was found that over 80 percent were pine shives. Extended studies were undertaken to determine the source of these pine shives. (14) Results of these studies showed we were getting pine pulp to the bleach plant from the following two sources: (1) from mixing of pine and hardwood chips ahead of the batch digester and (2) from mixing of hardwood and pine pulp when switching from hardwood to pine on our brownstock washers. Opportunities for mixing of pine and hardwood chips and pulp are high in our case since no separate wood handling or pulp processing line exists for hardwood.

Subsequent testing showed both sources of pine contamination resulted in high K-No. pine pulp (25 to 48 K-No.) being sent to the bleach plant. Laboratory digester cooks on mixtures of pine and hardwood showed that the hardwood was overcooked and the pine undercooked. Also the pine pulp which mixed with the hardwood on brownstock washer transitions was at a high K-No. since the pine was a relatively high yield pulp for unbleached linerboard production.

High K-No. pine pulp resulting from either of the above causes would be more difficult to bleach and could cause shive problems. This is particularly true in our case since the hardwood stock for our bleach plant is hot stock refined and screened with no rejects being removed from the brownstock system. Therefore, none of the high K-No. pine will be removed prior to the bleach plant.

A flowsheet of our brownstock preparation system is shown in Figure 3. We are continuing to use this flowsheet for processing our hardwood pulp prior to bleaching. To minimize the problems this flowsheet presents we are taking all steps possible to prevent mixing of pine and hardwood chips ahead of the digester as well as pine contamination during transitions from pine to hardwood on the brownstock washers.

It should be noted that it is very unusual for a closed brownstock system utilizing hot stock refining and screening to be used in preparing bleachable grade hardwood pulp. Normally there is a concentrated brownstock stream rejected from knotters, screens, centrifugal cleaners, or combinations of these cleaning devices. This reject stream is normally recooked or used in a product requiring less cleanliness. In Chesapeake's case there is no reject stream and all rejects are refined and sent to the bleach plant. Also it is normal in a bleached mill to have chip screens to remove oversize chips as well as fines. Chesapeake has no chip screens. The ability of the oxygen bleach system to produce a pulp which is low in dirt and shive levels from a 16 K-No. pulp processed by Chesapeake's flowsheet shows the large dirt and shive removal potential of the oxygen bleach system.

Other Important Bleach Pulp Properties—

Minimum problems have been experienced in obtaining acceptable brightness or dirt levels at design tonnage rates. Also bulk and opacity, which are very important properties to be maintained in bleached hardwood pulps, have been at acceptable levels. Table 2 compares the properties of pulp from the D/COD sequence to that of of 5 other Southern hardwood pulps. The comparison is made at a 0.5-percent cupriethylene diamine (CED) viscosity of 10 to 10.5 and a 300 CSF freeness level. Also data are given at two different levels of pine since pine content can have a large effect on the properties of the final pulp. As Table 2 shows, properties of the oxygen bleached pulp are very similar to those of the southern hardwood pulps bleached by conventional sequences.

Effluent Reduction Measurements

Initial laboratory results by CIL had indicated that the D/COD sequence should produce less total effluent color and BOD_5 than conventional chlorination sequences. Also the data indicated that effluent from the O-stage would be a low volume, low chloride concentration waste stream which, if recycled and fully recovered, would result in a significant further decrease in bleach plant effluent discharge. (15)

Combined Effluent Data—

Initial data gathered during the grant period indicated that combined BOD_5 and color levels, without O-stage recycle were about 13.5 and 42 kg/metric a.d. ton (27 and 84 lb/a.d. ton) respectively versus 17.7 and 100 kg/metric a.d. ton (35 and 200 lb/a.d. ton) for conventional chlorination sequences. A 23-percent reduction in BOD_5 and a 58-percent reduction in color was reported as typical effluent reductions to be expected from the D/COD sequence. (16) Subsequent testing of bleach plant effluents after the grant period did not confirm initial results. In fact this more reliable set

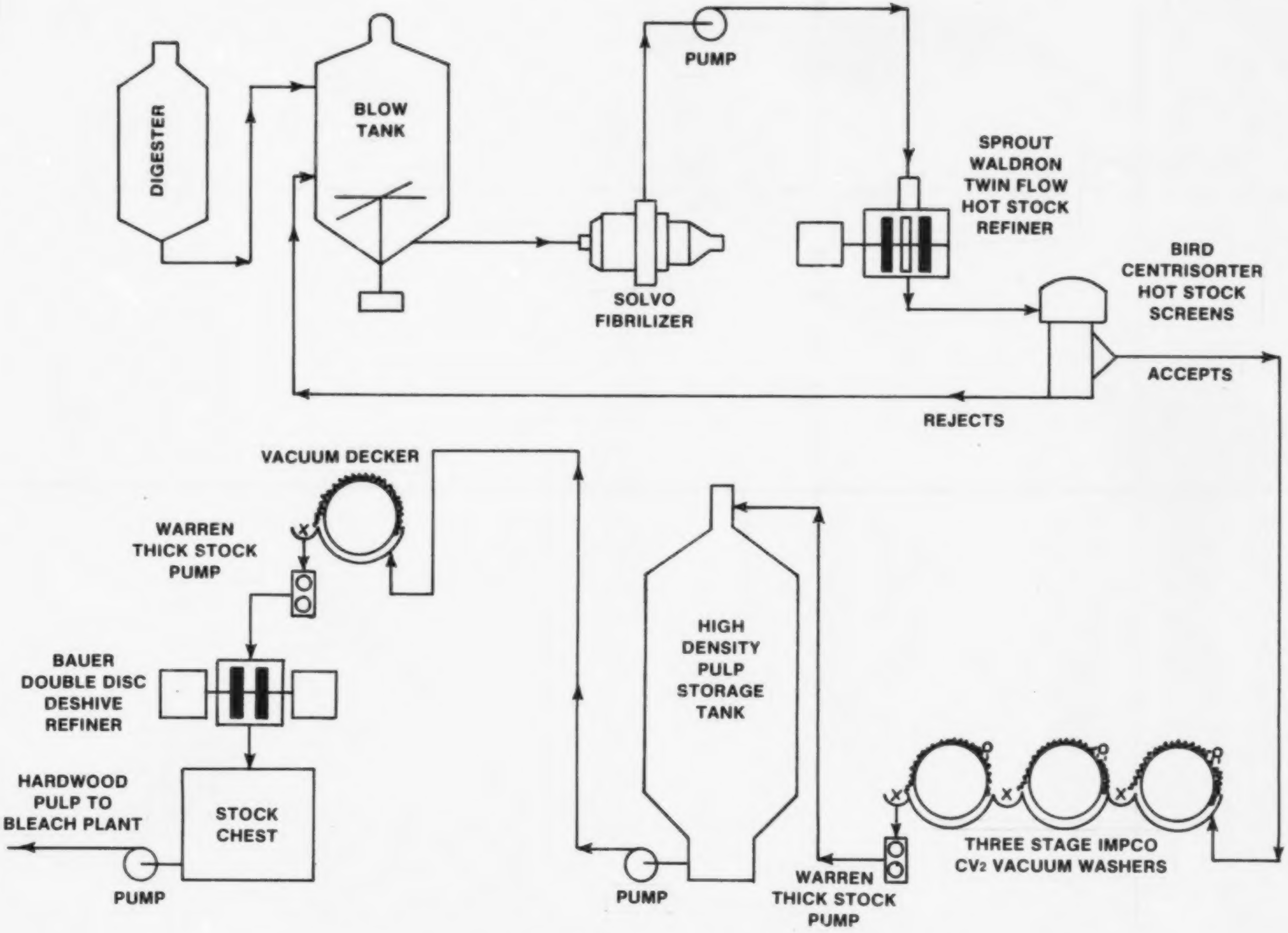
DIGESTER
BLOW TANK
PUMP
SOLVO FIBRILIZER
SPROUT WALDRON TWIN FLOW HOT STOCK REFINER
BIRD CENTRISORTER HOT STOCK SCREENS
ACCEPTS
REJECTS
VACUUM DECKER
WARREN THICK STOCK PUMP
BAUER DOUBLE DISC DESHIVE REFINER
HARDWOOD PULP TO BLEACH PLANT
STOCK CHEST
PUMP
HIGH DENSITY PULP STORAGE TANK
PUMP
WARREN THICK STOCK PUMP
THREE STAGE IMPCO CV2 VACUUM WASHERS

Figure 3. Brownstock flowsheet for bleachable grade hardwood pulp.

TABLE 2. PROPERTIES OF OXYGEN AND CONVENTIONALLY BLEACHED PULPS AT 300 CSF

Mill	Burst (factor)	Breaking length (meters)	Tear (factor)	Bulk cm^3/g	Opacity (%)	Brightness units (GE)	Shives (number)	Dirt (cm^2/m^2)	0.5% CED Viscosity (centipoise)	Hardwood (%)
A	43	6800	57	1.47	74.3	85.6*	—	—	10.0	95
B	46	6900	55	1.46	75.0	86.1*	0	.2	10.3	90
C	51	7050	95	1.47	—	87.0*	1-2	.6	10.4	85
D	47	7000	95	1.47	71.0	84.2*	5-6	1.1	11.1	70
E	49	7200	88	1.47	75.8	84.3*	—	—	10.5	61
F	—	—	—	—	—		1-4	.7	—	—
G	—	—	—	—	—		7—8	.7	—	—
H	—	—	—	—	—		1—2	1.6	—	—
Mean ±	47	6900	78	1.47	74.0	89.0	2½-3½	.8	10.3	87±
	± 3	± 150	± 20	± .005	± 2.1			± .5	± .2	19
Chesapeake pulp	43	6600	85	1.49	75.5	88+	Less than 3	Less than 1	10.5	95
CCV pulp with pine added	47½	6675	90	1.50	72.7	88+	Less than 3	Less than 1	11.0	80

*Samples had naturally aged. Estimated original brightness several units higher.

of data indicates the total BOD_5 discharge from the bleach plant was higher than conventional sequences and that color reduction was about 36 percent rather than 58 percent. (17) In analyzing these two sets of measurements it was concluded that the combined effluent flow meter was in error when the initial set of data was collected. This resulted in the generation of lower initial effluent discharge levels.

Table 3 compares the combined effluent data from the D/COD sequence to the original CIL laboratory-based projections, and to levels from conventional sequences. Flow, BOD_5, and color data for conventional sequences is based on a 1974—1975 survey of 23 mills by the NCASI. (18) Data on chlorides is taken from literature references and available NCASI mill data. (7, 18, 19) Table 3 shows that the three-stage D/COD sequence produces a combined effluent low in volume and color. However, BOD_5 levels are higher than these from conventional sequences. Also, BOD_5, color and chlorides were all considerably higher than those projected from CIL's original laboratory studies.

It should be emphasized that the BOD_5, color, and chloride levels from the D/COD sequence contained in Table 3 result from bleaching hardwood pulps at about 16 K-No. It has been our experience that brownstock K-No's. for conventional sequences are normally in the 10 to 12 range. Lower color BOD_5 and chloride levels would be expected in bleaching pulps at these lower unbleached K-No's.

TABLE 3. COMPARISON OF COMBINED EFFLUENT PROPERTIES

Sequence	Flow*	BOD_5#	Color#	Chlorides #
D/COD (mill data)	36,700 (8,800)	23.8 (47.6)	70.4 (140.9)	37.3 (74.5)
D/DOD (CIL laboratory projections)	——	16 (32)	52.6 (105.2)	25.3 (50.5)
Conventional 4—6 stage bleach sequences	82,200-116,840 (19,700-28,000)	17—20.5 (34—41)	105—117.5 (210—235)	29—43 (58—86)

*Flows are in L/a.d. metric ton (gal/a.d. ton)

#BOD_5, color, and chlorides are in kg/a.d. metric ton (lb/a.d. ton)

Effect of Recycling O-Stage Effluent—

Table 4 shows stage by stage data from the D/COD sequence. This data was taken in 1976 after installation of the D/C washer. This data shows that the O-stage effluent is a relatively low volume stream which contains about 54 percent of the total BOD_5 and 45 percent of the total color. However, it only contains 14 percent of the chlorides. With O-stage recycle, assuming complete destruction of organics, BOD_5 could be reduced to 11 kg/metric a.d. ton (22/a.d. ton) and color to 38.5 kg/metric a.d. ton (77 lb/a.d. ton). This would result in a 35 to 46-percent reduction in BOD_5 and a 63 to 67-percent reduction in color when compared to levels from conventional sequences (Table 3).

The building and operation of a pilot plant to reduce chlorides in white liquor so as to maximize return of oxygen stage effluent to chemical recovery, was an original objective of the EPA grant. However, by the time Chesapeake had sufficient technical manpower to begin this part of the project two significant technical advances had occurred which made completion of this objective questionable. These technical advances included (1) the development by CIL of the O-D/CED sequence to the point they were ready to offer it to the pulp and paper industry (20) and (2) the commercialization of the Reeves-Rapson salt recovery process by Erco Envirotech. (21) The development of the O-D/CED sequence would allow recycle of O-stage effluents without need for a salt removal system and the commercialization of the Erco "Salt Recovery Process" meant a salt removal system was now available to the pulp and paper industry.

TABLE 4. D/COD STAGE-BY-STAGE EFFLUENT DATA

Source	Flow*	BOD_5 #	Color #	Sodium #	Chlorides #
D/C	16,734 (4,010)	7.65 (15.3)	34.5 (69.0)	9.2 (18.4)	26.5 (53.0)
O	6,552 (1,570)	12.8 (25.6)	32 (64.0)	12.8 (25.5)	5.3 (10.5)
D	8,367 (2,005)	3.35 (6.7)	3.9 (7.9)	5.7 (11.4)	5.2 (10.3)
Process water calculated†	5,154 (1,235)	0	0	0.095 (1.9)	0.35 (0.7)
Total	36,806 (8,820)	23.8 (47.6)	70.5 (140.9)	28.6 (57.2)	37.3 (74.5)
O-stage (percent of total)	17.8	53.7	45.4	44.6	14.1
Effluent levels with O-stage recycle	——	11.0 (22.0)	38.5 (76.9)	——	——
Percent reduction from conventional sequence levels (see Table 3)	——	35-46	63-67	——	——

*Flows are given in L/a.d. metric ton (gal/a.d. ton)

#BOD_5, color, sodium, and chlorides are given in kg/a.d. metric ton (lb/a.d. ton).

†Pulp seal water, cooling tower blowdown, etc., from SOP and bleach plant areas.

In light of these developments it did not appear that efforts by Chesapeake to develop a crystallizer in conjunction with an oxygen bleaching system was necessary. However, sufficient data was collected to calculate the equilibrium white liquor sodium chloride concentrations which would result from recycling the O-stage effluent from a D/COD sequence. These 1976 data indicated that recycling O-stage effluent would have raised Chesapeake's white liquor sodium chloride levels from about 6 to about 17 g/L. Bleached pulp production from 1976 averaged about 227 metric a.d. tons/day (250 a.d. tons/day) with an unbleached kraft production of about 952 metric a.d. tons/day (1050 tons/day).

It is interesting to note that the projected 17 g/L salt level is near the apparent design level of 12 g/L for the mill at Thunderbay, Ontario which has procured the closed bleach plant system using the Erco Salt Recovery Process. (22) If this is in fact the design salt concentration level, it is unlikely that the development of a crystallization process to lower salt levels from 17 to 12 g/L would have been justified in Chesapeake's particular case.

OXYGEN PLANT AND DISTRIBUTION SYSTEM

The oxygen plant represents no technological advance. However, its installation by Chesapeake does represent a drastic departure from the traditional conservatism of the pulp and paper industry. Also, the oxygen plant is a key part, technically and economically, of the development of the various oxygen applications. The plant was purchased from Burdett Oxygen Company of Norristown, Pennsylvania, and supplied by Superior Air Products of Sayreville, New Jersey.

Oxygen Plant Description*

The cryogenic process is used to produce high-purity-oxygen from air. Air is compressed in a large four-stage Joy compressor to approximately 1,035 kPa (150 psig). After each compression stage the air is cooled by indirect heat exchange with water from a Marley cross-flow cooling tower. After the 4th stage the air is further cooled by indirect heat exchange with water chilled by direct contact with dry waste nitrogen.

Carbon dioxide and water are then removed by adsorption on molecular sieves in a two-bed dryer system. Thereafter the process air is cooled in a main heat exchanger by outgoing oxygen and waste nitrogen gas, then passed through a silica gel acetylene adsorber before entering the lower section of a double distillation column. Liquid oxygen from the bottom of the upper column feeds to a 106,000 L (28,000 gal) liquid storage tank at 172 kPa (25 psig). It can be vaporized if needed and fed to the gaseous oxygen distribution system. Normally, excess liquid oxygen is sold to offset the already low cost of oxygen production.

The oxygen plant is designed to produce 91 metric tons/day (100 tons/day) of 99.5 percent purity oxygen gas at 172 kPa (25 psig) and ambient temperature or up to 9.1 metric tons/day (10 tons/day) of liquid plus 81.6 metric tons/day (90 tons/day) of gas at a consumption of 1865 to 1940 kW (2500 to 2600 hp).

Oxygen Distribution System Description

The oxygen distribution system carries gaseous oxygen to different applications at the required rate and pressure. The mild steel pipeline, degreased and dried before startup, receives oxygen at 172 kPa (25 psig) from the oxygen plant (Figure 4) or, on a short-term basis, from liquid storage by an automatic steam vaporizer. Sufficient backup liquid storage capacity is maintained to allow an oxygen plant shutdown of 1 to 2 days.

A pressure of 172 kPa (25 psig) is sufficient for the wastewater treatment plant and the proposed lime kiln consumption. A centrifugal wet Nash compressor provides 483 kPa (70 psig) for black liquor oxidation and a two-stage Sulzer reciprocating oxygen compressor provides 1,034 kPa (150 psig) at the bleach plant. A 11,355 L (3000 gal) liquid-oxygen storage and air vaporizer backs up the bleach plant compressor and provides large quantities of oxygen rapidly for startup of the bleach plant reactor.

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*U.S. Patent 3,070,966

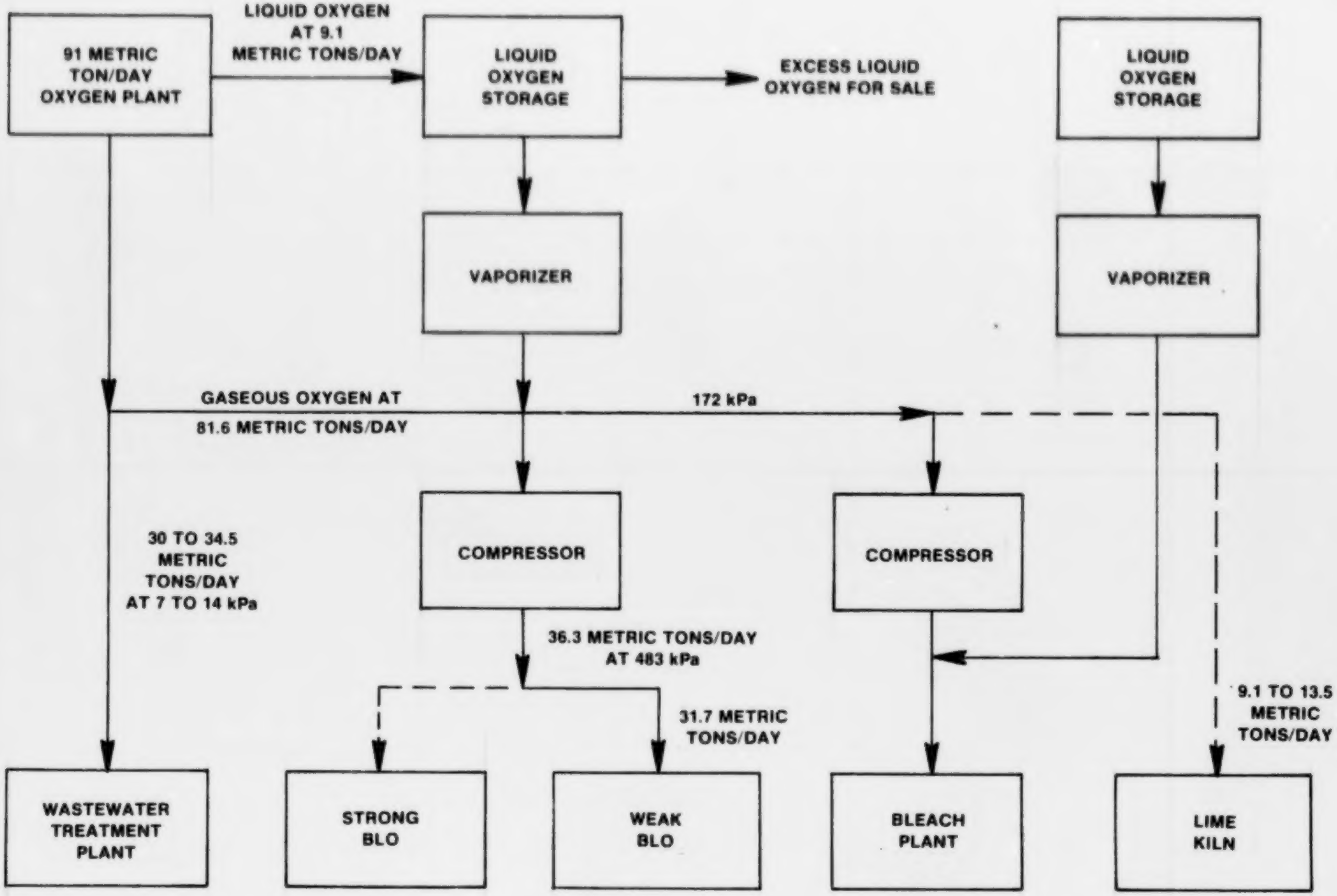
91 METRIC TON/DAY OXYGEN PLANT
LIQUID OXYGEN AT 9.1 METRIC TONS/DAY
LIQUID OXYGEN STORAGE
EXCESS LIQUID OXYGEN FOR SALE
LIQUID OXYGEN STORAGE
VAPORIZER
VAPORIZER
GASEOUS OXYGEN AT 81.6 METRIC TONS/DAY
172 kPa
30 TO 34.5 METRIC TONS/DAY AT 7 TO 14 kPa
COMPRESSOR
COMPRESSOR
36.3 METRIC TONS/DAY AT 483 kPa
31.7 METRIC TONS/DAY
9.1 TO 13.5 METRIC TONS/DAY
WASTEWATER TREATMENT PLANT
STRONG BLO
WEAK BLO
BLEACH PLANT
LIME KILN

Figure 4. Oxygen distribution system.

Operating Problems

Most of the problems with the oxygen plant have been mechanical and have been associated with the large air compressor's related equipment and automatic controls. However, several process problems did result. One was the high noise level from the oxygen plant which was located 100 m (328 ft) from the nearest home. Additional silencers have solved this problem. A second problem was the lack of turn down capability of the oxygen plant. This is no longer a problem since all of the plant capacity is now being used. However, turn down capability of a cryogenic oxygen plant can be very limited if the plant is not designed with turn down in mind. This can result in high oxygen power costs at lower production levels. A third problem occurred in the dryer section which removes carbon dioxide and water prior to the cooling of air in the main heat exchanger. A faulty condensate trap, removing water from initial cooling of the air ahead of the dryer section, allowed liquid water to carryover into the molecular sieve dryers. This resulted in water being carried through the dryer section and caused "icing" and plugging within the heat exchangers following the dryer section. Also it caused problems with regeneration of the dryers as well as a reduction in the service life of the molecular sieves. Since modifications to the trap have been made this problem has been eliminated.

Startup problems with the oxygen distribution system were also few except for some problems in selecting materials of construction for the oxygen distribution piping and in cleaning the distribution piping for oxygen transport. In two cases, carbon steel piping was replaced with stainless steel piping prior to use of oxygen. In the first case, the oxygen line from the liquid storage tank into the bleach plant was changed from carbon to stainless to provide better protection in the bleach plant against possible pipeline fires. In the second case, the oxygen pipeline downstream of the wet compressor for black liquor oxidation was changed from carbon to stainless steel to provide better corrosion protection against the wet oxygen gas.

Problems with cleaning the oxygen distribution piping resulted primarily from two things First, the 549-m (1800-ft) section of pipeline from the oxygen plant into the mill area was carbon steel. Carbon steel piping requires a more elaborate cleaning procedure than stainless. Second, while installing the pipeline, there were no provisions made for cleaning the piping. Drains for removing the chemical cleaning solutions were not provided. Also, the piping had a number of low points that could not be easily drained. As a result, cleaning and drying the piping before the introduction of oxygen was made much more difficult.

Making certain the gas distribution system was essentially leak proof also required a large amount of labor and time. Most of these leaks were at threaded joints, for example, the caps on the drains installed for cleaning purposes. Aside from the above startup problems, we have experienced no serious problems to date with conveying oxygen in the distribution piping.

Operating Results

The oxygen plant is consistently producing the design quantity of liquid and gaseous oxygen except during extremely hot and humid days. Also, the design oxygen purity level of 99.5 percent is being met.

The oxygen plant is essentially automated and requires little operator attention while running. Defrosting the oxygen plant and starting the plant up from a warm condition, however, does require operator and supervisory attention. Cold startups, on the other hand, are less complicated and can now be handled by the operators with little supervisor attention. Defrosting of the oxygen plant is recommended once every 12 months to remove buildup of water and CO_2 in the heat exchangers, piping, etc. Other operating characteristics of the oxygen plant as supplied by SUPAIRCO are shown in Table 5. Operating experience is in agreement with the defrost time and cold and warm start times indicated in this table. Defrost intervals have also been extended to 12 months or greater since the dryer section problem, previously discussed, has been solved.

TABLE 5. OPERATING CHARACTERISTICS OF THE OXYGEN PLANT

Characteristic	Time
Defrost Time	24 hours
Warm start time	24 hours
Cold start time (short shutdown)	1—2 hours
Continuous operation time without defrost	12 months

WASTEWATER TREATMENT PLANT

The process design of the WWTP facility was based on a cooperative 8-month pilot plant study between Chesapeake, Union Carbide, and Hydroscience, Inc., wastewater consultants of Westwood, New Jersey. (23) The pilot plant was supervised and operated by Chesapeake at the West Point mill. A description of these pilot plant studies is included in Appendix C.

Claimed advantages of this system over conventional systems are (1) less land space, (2) higher oxygen transfer rate to the liquid waste per unit of energy input (3) 90-percent oxygen utilization, (4) fast response to fluctuating organic waste loading, (5) virtual elimination of odor, (6) less sludge production, (7) exceptionally high sludge settling rates and compaction densities, and (8) easier sludge dewatering.

Waste Treatment Plant Description

The full-scale wastewater treatment plant was mechanically engineered and installed by Rust Engineering. The process flowsheet, as well as the equipment used in the wastewater treatment plant, are similar to several activated sludge plants in the pulp and paper industry with the exception of the aeration basins (Figure 5). The aeration basins, or bioreactors, are covered to ensure a high utilization efficiency of the oxygen by the activated sludge. A 90 percent utilization efficiency of oxygen is expected by use of these covered oxygen reactors.

The bioreactors consist of a large covered concrete tank divided into two parallel, three-stage systems by a central mid-feather. Conventional surface aerators provide the necessary oxygen transfer. Oxygen feed gas is piped into the gas space above the first stages of the reactor. Wastewaters and recycled activated sludge are also fed into the first stages to provide cocurrent flow.

The oxygen feed control to the bioreactors is automatically regulated to maintain a set gas pressure of a few inches of water in the reactors. In order to maintain the necessary purity of oxygen above the liquid phase in each stage, some oxygen gas in vented from the the last stage along with the CO_2 generated from biological stabilization of the wastewaters. The amount of oxygen to vent is determined by monitoring the vent gas flow and the vent gas composition with an oxygen analyzer.

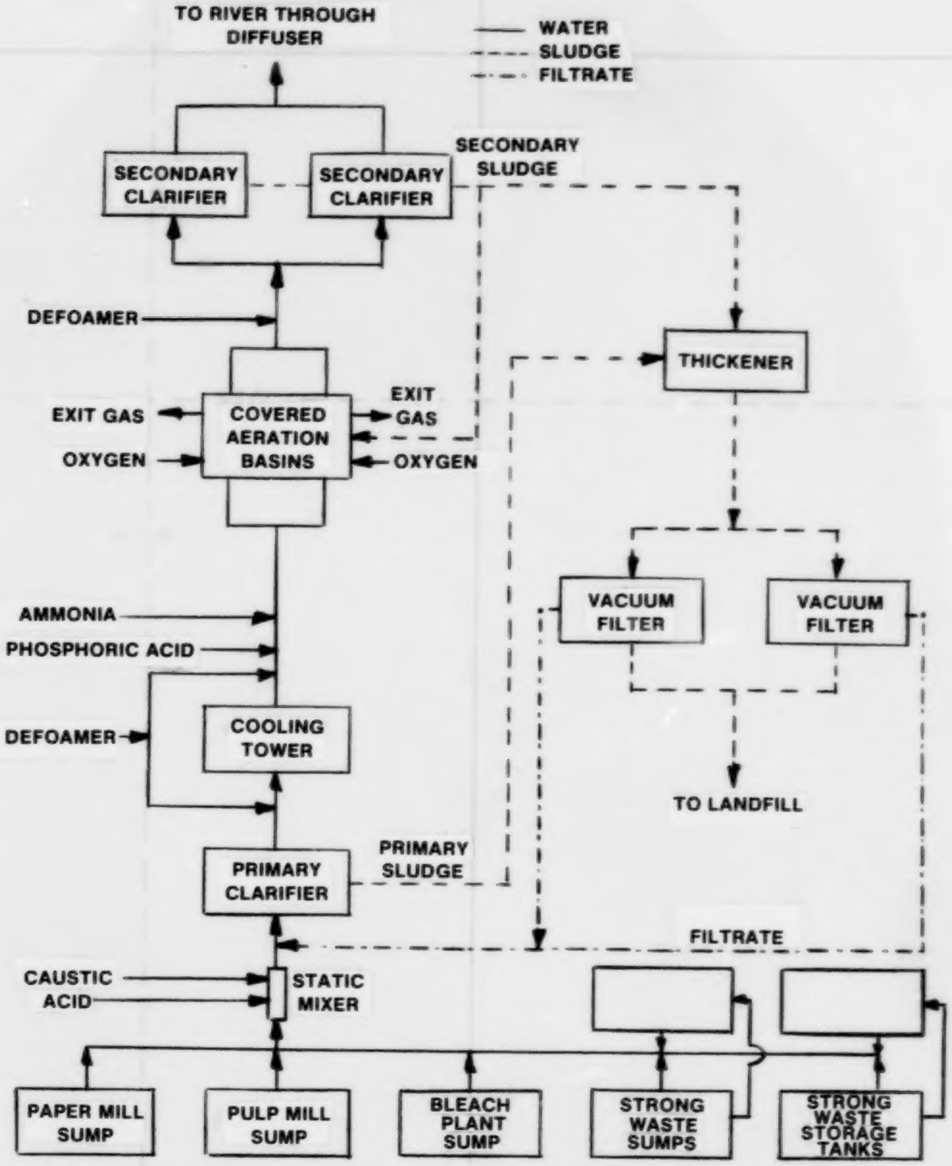

Figure 5. Wastewater treatment plant flowsheet.

Explosive mixtures of hydrocarbons (such as turpentine) and oxygen can occur in the gas space above the liquid level in the reactors. Several precautions were taken to prevent this. First, no combustible materials are installed under the reactor cover and there are no electrical circuits or components in the gas space. The aerator drives are located outside on the roof of the reactor and the agitators are driven through water seals to eliminate any metal-to-metal contact inside the tank covers. Second, any potentially dangerous accumulation of hydrocarbons in the gas space above the liquid level is sensed by a hydrocarbon monitor purge system in the gases from the first and last stages of the reactor. When the level of hydrocarbons reaches 25 percent of the lower explosive limit oxygen flow automatically shuts off and an air blower purges the hydrocarbons from the system. If the hydrocarbon level continues to rise and reaches 50 percent of the lower explosive limit, the aerators also automatically shut down.

The oxygen reactors are further protected from excessive pressure or vacuums by a series of pressure-vacuum breakers on each stage. Also, the water seals around the aerator shafts serve as pressure-vacuum breakers.

Operating Problems

Operating problems with the collection system and WWTP were numerous but many were confined to parts of the system which were of conventional design. Plugging of lift pumps in the four initial mill sumps was a continuous problem until one of the two pumps at each sump was replaced with a pump that could pass up to 7 mm (2-¾ in.) diameter particles. At the WWTP, sludge line plugging occurred initially but was reduced when butterfly control and block valves were replaced with V-ball and gate valves, respectively. Also, all check valves in sludge lines were removed. Scum line plugging and scum pumping was made much more difficult at Chesapeake by additional floating plastic and styrofoam material from the mill post-consumer waste fiber system. To minimize the problem the scum is now discharged directly to a perforated dumpster box which retains most of the plastic material; however, some of this material overflows the primary clarifier to the cooling tower cold well where it floats again. At present, we remove this floating material manually. No good solution has been found to this problem.

Operation of the cooling tower has been very satisfactory. Plugging occurred in the metering orifices/diffusers in the distribution plate on top of the tower and slime built up in the tower fill. However, these problems are controllable. Foaming in the cooling tower and drift loss were expected. Neither have been a serious problem.

Problems operating the bioreactors have also been minimal. Prior to startup, finding and stopping all gas leaks in the concrete basin was a problem. Then when heavy organic loads were first fed to the plant, primarily due to black liquor spillages, the aerator motors overloaded. Heavy foam in the gas space was suspected. Further study showed the foam increased the level in the reactor by partially blocking flow between the reactor and secondary clarifiers. This increased submergence and load on the fixed level aerators. The problem was solved by adding defoamer. Defoamer is added continuously to the influent to the cooling tower and, during black liquor spillages, to the inlet and exit flows of the bioreactors. Solving problems in the oxygen reactors, such as the one described above, is hampered by the fact that there is no way to view the interior.

On two occasions over a period of 6 to 7 months bad spillages of turpentine into the mill sewer system have caused the hydrocarbon analyzer to activate the air-purge system on the covered bioreactor. When this air-purge system is in operation, the amount of organic wastes that can be stabilized by the treatment plant is substantially reduced since the oxygen partial pressure drops sharply in the gas space above the liquid level. To date, the hydrocarbon levels reached have not shut down the aerators. Turpentine spills to the sewer have essentially stopped after changing in-plant equipment and procedures. Also it was determined that at startup and for sometime thereafter, the set point of the hydrocarbon analyzer was mistakenly adjusted to 12.5 percent of the lower explosive limit (LEL) for air-purge blower startup. This was one-half of the normal set point. No purge of the reactors has been necessary since in-plant turpentine spills have been controlled and the hydrocarbon analyzer reset.

The peripheral feed, peripheral overflow secondary clarifiers have functioned well mechanically. The sludge is removed from the clarifiers by bottom sweeping sludge withdrawal mechanisms which operate by hydrostatic head differential. These mechanisms have successfully removed sludge nearly 3 percent in concentration without plugging. Foam on the inlet raceway of the secondary clarifiers has been a problem whenever the treatment plant receives a very high organic load due to black liquor. Foam has developed in the raceway to the extent that it has overflowed the outside clarifier walls as well as into the effluent launder. A partial solution to this problem has been the addition of defoamer in the UNOX reactor effluent at the combined liquor weir, before the secondary clarifiers and an increase in the height of the feed launders. Foam problems still exist to some degree in the influent launders. Chesapeake has no recommendations to offer as a certain solution to the problem except possibly a large free surface stilling well between the bioreactor and secondary clarifiers. Also the transfer of the mixed liquor from the bioreactors should be accomplished with a minimum of turbulence and chance of air entrainment. The need for the above precaution is especially necessary in an oxygen WWTP because of larger amounts of dissolved carbon dioxide in the mixed liquor than with conventional air systems.

Achieving design secondary clarifier underflow and bioreactor mixed liquor suspended solids levels was initially a difficult problem which still exists to some degree. The primary cause for both of these problems has been the growth of filamentous organisms in the bacterial population. Extensive pilot and mill scale trials have shown that these filamentous organisms can be controlled in our system by the addition of hydrogen peroxide to either the biological sludge being recycled to the bioreactors or to the feed to the primary clarifiers. (24) The control mechanism proposed is the oxidation of sulfides occurring in the WWTP influent.The principal filamentous organisms identified as present in Chesapeake's bacterial population use sulfide as a food source. Elimination of these sulfides should theoretically "starve" these organisms and minimize the filamentous problems. Hydrogen peroxide has been used successfully on three occasions since 1975 to control filamentous growth. This has allowed the plant to operate at design bacterial populations in the bioreactors (5500 mg/L) and near design secondary clarifier underflow solids levels (2.5 percent). It has also helped to reduce effluent suspended solids levels.

Satisfactory solutions to the problems previously discussed has resulted in consistent WWTP plant operation during the past several years. Problems still exist in two areas, however. One is higher and more erratic suspended solids levels in the final clarified effluent than expected. This we feel may be improved by a redesign of the piping between the bioreactors and secondary clarifiers to minimize turbulence and air entrainment. Also a large free surface stilling well between the bioreactors and secondary clarifiers would be a definite improvement. As previously discussed these steps would also minimize foam. Particular attention to design of this part of the oxygen WWTP should be taken since the higher carbon dioxide levels in the mixed liquor discharged from the oxygen bioreactors could cause foaming as well as clarification problems in secondary clarifiers.

The second problem which exists is the age-old sludge disposal problem. Less biological sludge, which dewaters reasonably well, was obtained from the oxygen WWTP. However, landfilling of mixtures of primary and secondary sludge of high organic content (80 to 90 percent) has been a continuous and difficult learning experience. Present operation of our landfills are markedly improved. However, under almost any circumstances creating a landfill that is structurally stable, is free from potential leaching problems and is insect and odor free is very difficult. Alternate solutions to landfilling are under continuing study.

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Operating Results

Operating results of the fullscale WWTP system after the solution of initial operating problems have been good and have demonstrated the claimed advantages of the high purity oxygen system. These results compare favorably with pilot plant and design data at equivalent BOD_5 loading rates with the exception that higher effluent suspended solids levels have been experienced in the full scale plant. Suspended solids levels in the past three years have averaged between 89 to 98 mg/L compared to a 50 mg/L design level. Table 6 shows average operating characteristics of the WWTP in May and June of 1977. (25) These results are typical of the WWTP operation at design BOD_5 loading rates. Typical sludge settling and compaction characteristics with and without filamentous organisms present are shown in Table 7. (26) As previously mentioned filamentous organisms had a detrimental effect on both of these sludge characteristics.

BLACK LIQUOR OXIDATION SYSTEM

While there are many BLO systems which use air as the oxidant, Chesapeake decided on pipeline injection of oxygen because of its potential cost advantages, both capital and operational. The design of the BLO system was based on both bench and pilot scale studies by Chesapeake's Research Department. (27) These studies provided information on reactor design and control, economic optimization, and side effects.

TABLE 6. TYPICAL OPERATING CHARACTERISTICS OF WWTP

Parameter	Operating characteristic	
Influent to bioreactor		
Flow, L/s (mgd)	525	(12)
BOD_5, mg/L (lb/day)	425	(42,580)
Total suspended solids, mg/L	100	(10,016)
Average biomass loading, kg BOD_5/day/kg MLVSS*	0.66	
Final effluent		
BOD_5, mg/L (lb/day)	40	(4050)
Total suspended solids, mg/L (lb/day)	79	(7800)
BOD_5 removal efficiency (percent)	90.5	
Solids production (kg total suspended solids per kg of BOD_5 removed)	0.43	
Oxygen utilization efficiency (percent)	94.2	

*Mixed liquor volatile suspended solids

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Maximizing the oxygen utilization efficiency was the chief objective in reactor design. The studies at Chesapeake concluded that the single most influential variable on the OUE was the shear rate, not the Reynolds Number as had popularly been assumed, of the liquid phase in the reactor. To assure a high OUE the BLO system was designed for a shear rate of -1.5×10^6 min $^{-1}$. By maintaining this shear rate, oxidation can be achieved without the use of a venturi or any kind of static mixer.

Description of the BLO System

Equipment in the BLO system consists of a Nash compressor that feeds oxygen gas to two oxygen spargers in the liquor line from the weak black liquor storage tank to the multiple-effect evaporators (Figure 6). The oxygen sparger now in use has 372 holes, 1.6 mm (1/16-in.) in diameter. The second sparger serves as a spare. These spargers are followed by 30.5 m (100 ft) of 203 mm (8-in.) pipe which provides the proper shear rate and retention time for oxidation to occur. There is another 83.8 m (275 ft) of 254 mm (10 in.) pipe to the evaporators which provides additional retention time.

The oxygen feed to the sparger is controlled by measuring the black liquor flow rate and feeding oxygen gas at a combined ratio to the liquor flow. This ratio is automatically adjusted to provided a set residual sodium sulfide concentration in the oxidized liquor which is measured continuously by a silver billet tamal reference electrode pair.

Operating Problems

Control of weak liquor flow to the three sets of evaporators was a significant startup problem. Relocation of some flow control and safety valves has resolved this problem.

TABLE 7. TYPICAL BIOLOGICAL CHARACTERISTICS OF WWTP

	Operating characteristic	
Parameter	With filamentous organisms present	With filamentous organisms present
Sludge volume index	90	50
Initial sludge settling velocity meters per hour (ft per hour)	1.07 (3.5)	1.68 (5.5)
Secondary clarifier underflow cencentrations (percent)	1.6	2.5
Mixed liquor total suspended solids levels (mg/L)	4000	4500

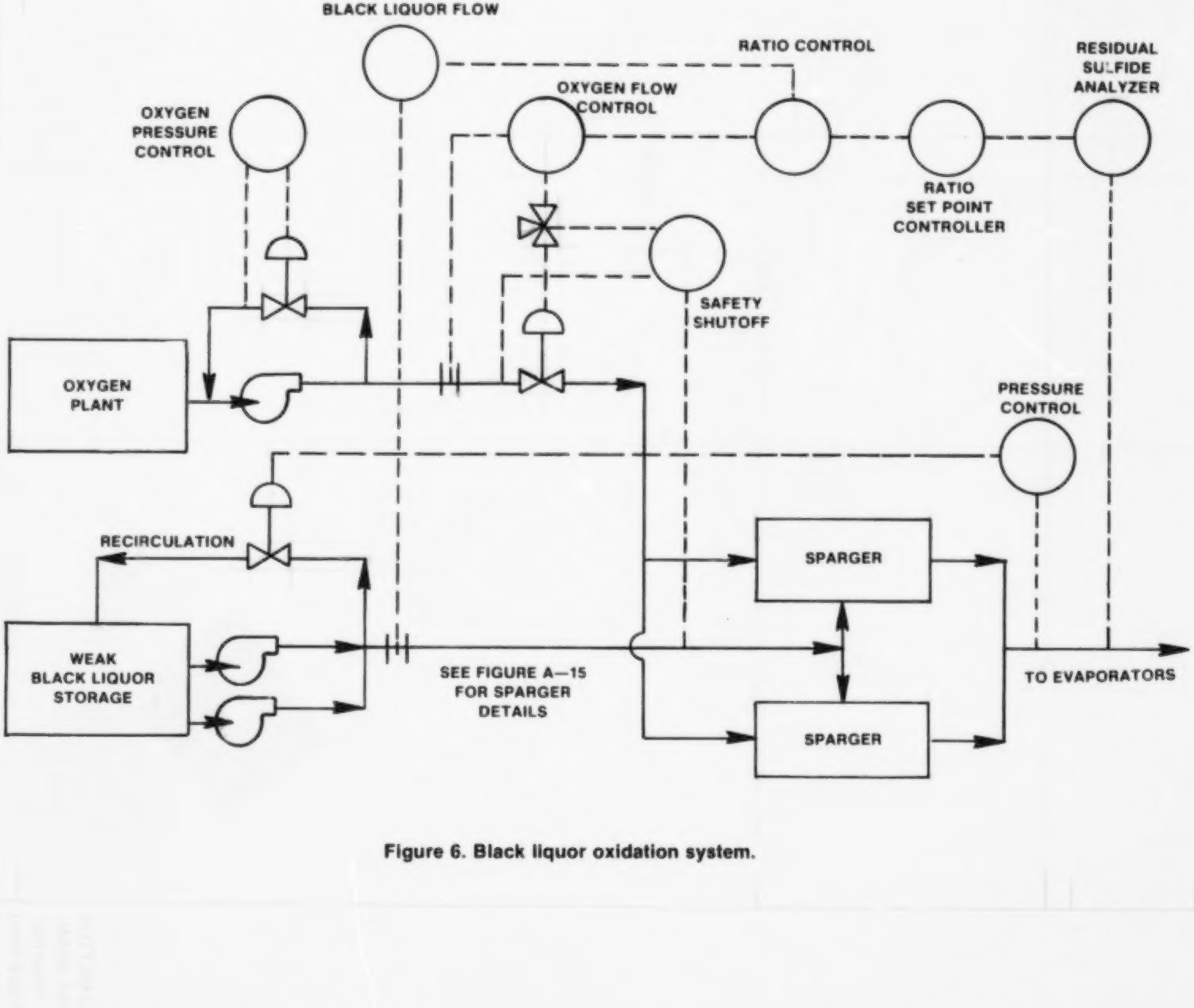

Figure 6. Black liquor oxidation system.

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Black liquor pressure control also presented a problem. In the full scale operation, the reactor pressure drop was significantly lower than was indicated from the pilot studies. This resulted in the oversizing of pumps and valves, ultimately creating a pressure control problem. By valve relocation and trimming the pump impellers we were able to reduce and control the liquor pressure as well as reduce the power consumption. The weak black liquor flow can now be controlled with steady flow conditions while maintaining a pressure at least 13.8 kPa (2 psi) below that of the oxygen. This pressure differential is necessary to prevent contamination of the oxygen distribution system with black liquor.

The life of the electrodes has been limited because of tip attack by the alkaline liquor. The electrodes were being replaced every 2 to 3 weeks, but it was found that by covering the electrodes with plastic "heat shrink" tubing, the life of the electrodes was extended to 2 to 3 months.

Oxygen utilization efficiencies are in the 75 to 80 percent range whereas over 90 percent was expected. Additional pilot plant studies, after full-scale BLO operation, were undertaken to attempt to define if OUE could be increased by further increasing shear rate. A high shear in-line Kenics mixer was used in addition to the pilot in-pipeline reactor. Results of these tests indicated no improvement in OUE. Therefore, a 75 to 80 percent OUE appears to be the maximum expected from an oxygen BLO system at an inlet sulfide level of 7 to 7.5 g/L. However, higher OUE can be obtained at higher inlet sulfide levels.

Operating Results

Typical operating results of the BLO system are included in Table 8. Conversion of the sodium sulfide to sodium thiosulfate is easily controllable at levels of more than 98 percent.

Reversion of only 2 percent (on a weak liquor basis) has been observed through the evaporators and another 2 percent while in heavy liquor storage. Measurement of odor level in the evaporator area has not been made but, subjectively, the evaporator operators have reported lower detectable odor in the work area. The exothermic oxidation reaction increases the black liquor temperature about 11 °C (20 °F) which helps increase the capacity of the evaporators and decrease the amount of steam required.

Overall, the high-purity-oxygen BLO system on weak liquor has operated smoothly. It is easy to startup and shutdown. The evaporators actually run better with BLO. There has been no discernible foam, no added scaling of the evaporators, and no problem with evaporator vacuums at the surface condensers. It occupies no space other than the compressor and a 102-mm (4-in.) oxygen pipeline feeding the sparger pipe in the existing weak black liquor pipeline. This system has reduced odorous sulfur emissions from the mill by about 6 to 6.5 kg/a.d. metric ton (12 to 13 lb/a.d. ton).

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TABLE 8. FULL-SCALE BLO OPERATING DATA

Item	Amount	
Unoxidized black liquor		
Solids (percent)	15	
Temperature, °C (°F)	96	(205)
Sodium sulfide level, g/L	7.5	
Flow, L/s (gpm)	123	(1950)
Oxidized black liquor		
Temperature, °C (°F)	108	(226)
Sodium sulfide level, g/L	0.1	
Black liquor from evaporators		
Solids (percent)	50	
Sodium sulfide level, g/L	0.99	
Black liquor from heavy liquor storage		
Solids (percent)	50	
Sodium sulfide level, g/L	1.58	
Reactor shear rate (min^{-1})	-1.98 x 10^6	
Oxygen utilization efficiency (percent)*	76	

*Stoichiometric amount of oxygen to convert sodium sulfide to sodium thiosulfate divided by actual consumptions.

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92. Law Engineering Company's Site Investigation, August 17, 1972.

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115. Letter, November 3, 1975. K. R. Gilbreath (CCV) to Michael L. Meadows (SWCB) regarding Failure of Discharge Valve.

116. Letter, April 27, 1976. K. R. Gilbreath (CCV) to Wade E. Landford (SWCB) regarding Work To Be Done on Discharge Control Valve.

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APPENDIX A

DESCRIPTION OF MAJOR EQUIPMENT

This appendix describes the purpose, type, specific dimensions, and other characteristics of major equipment in the systems covered by this report. This information was drawn from drawings, purchase orders, operating manuals, maintenance and engineering files, consultation with operating people, and memoranda in Chesapeake's records. Detailed information with respect to pumps, valves, minor chemical storages, and most pipelines is not included in this appendix. However, the information is readily available at Chesapeake and will be provided upon request when feasible. Such information is voluminous and quite complex. Therefore, Chesapeake suggests that any person or group with a comprehensive interest in details of process equipment, control systems, and associated piping make arrangements to visit its West Point site.

IN-PROCESS EQUIPMENT CHANGED

A number of in-process changes were made by Chesapeake to reduce daily BOD_5 effluent from the oxygen bleaching system. This enabled Chesapeake to (a) complete a pilot plant study of the best technology from wastewater treatment and (b) to startup the bleach plant about 12 months before the WWTP was operating. In this manner Chesapeake could operate the bleach plant and remainder of the mill under discharge Certificate No. 1260, issued in 1960, until a decision could be made on treatment technology, such technology could be approved by the SWCB, and the approved technology could be installed by Chesapeake. Appendix C describes the results achieved by making the in-process changes. This appendix reports the important features of the equipment installed.

Multi-Stage Vacuum Washers

Function	To replace old screw press pulp washers with a more efficient system for recovering maximum inorganic and organic chemicals for re-use and steam generation, respectively. Also, to reduce organic chemical (BOD_5) loss to the sewer system.
No. of Stages	3 countercurrent
Type of washer	Stationary center valve vacuum
Drum size	Diameter—3.5 m (11 ft 6 in.)
	Length—7.3 m (24 ft)
Capacity	680.3 metric tons/day (750 tons/day)
Construction	Wire—304 L stainless steel
	Vat—steel
	Wire support—steel drum with stainless steel deck

Steam Stripper

Function	To remove volatile BOD_5 and odorous gases from evaporator condensates, turpentine decanter underflow and blow heat condensate (after installation of gas collection and odor reduction system). Parts of the stripper system are described in the following sections.

Auto-Strainer—

Function	To remove small wood chips and fiber from blow-heat condensate feed stream.
Type	Self-cleaning, automatic back wash (76 mm [3 in.] Model A)
Capacity	18.9 L/s (300 gpm) at 14.7 kPa (2.14 psi) pressure drop
Construction	Body—cast iron
	Strainer—304 stainless steel
Unit dimentions	Diameter—0.47 m (18½ in.)
	Height—0.97 M (39 in.)

Main Heat Exchanger—

Function	To heat the stripper feed stream by heat exchange with hot, stripped condensate leaving the bottom of the stripper column.
Type	Initial was a plate and coil exchanger which developed plugging and gasket problems as explained in Appendix C.
	Present exchanger is a spiral type
Dimensions	Diameter—0.89 m (35 in.)
	Length—2.03 m (6 ft 8 in.)
	Surface—98 m^2 (1055 ft^2)
Construction	Shell—steel, lined with 304 stainless steel

Stripping Column—

Function	To provide contact between heated condensate and turpentine underflow streams in order to strip or remove volatile BOD_5 and TRS compounds from liquid phase to gas phase.
Type	Vertical steam stripper—distillation column with sectional type valve trays; 14 trays with 42 valve elements per tray.
Dimensions	Diameter—1.37 m (54 in.)
	Height—17.4 m (57 ft)
Condensate flow	15.7 to 18.9 L/s (250 to 300 gpm)
Steam flow	5580 kg/hr (12,400 lb/hr) at 414 kPa (60 psig)

Construction	304 stainless steel

Reflux Condenser—

Function	To condense part of the gas stream leaving the stripper column and return condensate to the 10th tray of column.
Type	Horizontal shell and tube
Cooling medium	Saline river water
Dimensions	Diameter—0.51 m (20 in.) O.D.
	Length—4.47 m (14 ft 8 in.)
Capacity	5037 kg (11,194 lb) of vapor from stripper
Temperature	100.6°C (213°F) of reflux condensate
Reflux ratio	13 to 14/1
Construction	Tubes—316 stainless steel
	Shell—304 stainless steel

Secondary Condenser—

Function	To receive vapor from reflux condenser, condense and send condensate to turpentine decanter or to turpentine recovery system elsewhere in the mill.
Type	Horizontal shell and tube heat exchanger
Cooling medium	Saline river water
Dimensions	Diameter—0.51 m (20 in.) O.D.
	Length—2.9 m (9 ft 6 in.)
Capacity	2033 kg (4518 lb/hr)
Construction	Shell—304 stainless steel
	Tubes—XM33 (alloy between Crucible 26-1 and Hasteloy 273)

Decanter—

Function	To receive condensate from secondary condenser to decant turpentine when such condensates are not being returned to the mill turpentine recovery system.
Dimensions	Diameter—1.83 m (6 ft) O.D.
	Height—2.14 m (7 ft)
Construction	304 stainless steel

Scrubber—

Function	To receive exit gas from the reflux condenser or the secondary condenser to remove odorous gases when such exit gases are not going to strong gas line to be incinerated.
Type	Vertical wet scrubber packed with 1.9 cm (¾ in.) ceramic packing, 77 percent open area.
Scrubbing liquid	0.95 L/s (15 gpm) white liquor, returned to white liquor system
Vapor flow	6.1 m^3/min (216 acfm) to scrubber
Dimensions	Diameter—0.36 m (14 in.) Height—4.3 m (14 ft)
Construction	Body—steel Packing—ceramic

Surface Condensers

Function	To receive and condense vapor from last effect of multiple-effect evaporators, so as to maintain desired vacuum on last effect of system. There are four multiple effect evaporators, Sets A, B, C, and D.
Type	All four sets have long, vertical shell and tube heat exchangers
Dimensions	Shell—all sets 7.83 m (26 ft) high by 4.27 m (14 ft) diameter Tubes—Sets A, B, and D have 1520 tubes 32 mm (1¼ in.) O.D. by 18 BWG by 7.3 m (24 ft) long; set C has 1382 tubes of the same dimensions.
Heat transfer area	Sets A, B, and D have 1022 m^2 (11,000 ft^2); Set C has 903 m^2 (10,000 ft^2)
Fluids involved	Tube side—saline, screened river water Shell side—water vapor and condensate
Flow rates	Set A, 32940 kg/hr (73,000 lb/hr) vapor; 267.8 L/s (4250 gpm) cooling water and (11 psi) pressure drop across tubes. Set B, 31,590 kg/hr (70,200 lb/hr) vapor; 271 L/s (4300 gpm) cooling water and 69 kPa (10 psi) pressure drop across tubes. Set C, 29,295 kg/hr (65,100 lb/hr) vapor; 283 L/s (3780 gpm) cooling water and 69 kPa (10 psi) pressure drop across tubes. Set D, 34,126 kg/hr (75,835 lb/hr) vapor; 316.1 L/s (5018 gpm) cooling water and 56.7 kPa (8.23 psi) pressure drop across tubes.
Construction	Carbon steel for condensate and black liquor contact; 316 L stainless steel for saline river water contact.

Entrainment Separators

Function	To reduce or eliminate black liquor carryover into condensate or vapor side of surface condensers following last effect of the evaporators.
Type	Cyclone entrainment separator on A and C set evaporators; located after last effect and before surface condensers. Also surface condenser on each set has wrap-around centrifugal separator.
Construction	316 L stainless steel
B and D set evaporators	B set has parallel last effects—the new one installed having a much larger vapor body to slow vapor velocity and, in turn, to diminish liquor carryover to the vapor side of the surface condenser; likewise, the new D set of evaporators has a large vapor body with the last effect to achieve the same result. Also B & D sets have wrap-around centrifugal separators on surface condensers.

OXYGEN BLEACH PLANT

The oxygen bleach plant consists of three stages: D/C, O, and D, as indicated in the main body of this report. Equipment in each stage which serves a unique function is described in this section. Where appropriate, initial, interim, and present equipment are included.

D/C Stage

The first stage serves to mix the bleaching chemicals well with the incoming unbleached pulp, retain the mixture a fixed time at incoming feed temperature and wash the semi-bleached pulp until relatively free of residual chlorine.

Mixing—
The importance of good mixing of chemicals and pulp in bleaching cannot be over emphasized. Both high and low energy systems, with high and low maintenance costs, respectively, were involved in Chesapeake's D/C stage.

Initially installed:	Fiscalin mixer
Function	To mix chlorine dioxide solution with unbleached pulp, 30 second retention time.
Connected hp	56 kW (75 hp), 1800 rpm
Maintenance cost	Major
Construction	Carbon steel, FRP lined
Type	Series 400
Interim change:	Fiscalin mixer
Function	To mix a mixture of chlorine and chlorine dioxide solution with unbleached pulp, 30 second retention time.

Connected hp	45 kW (60 hp), 1200 rpm
Final	Fiscalin mixer Retained as standby about 2 months, disconnected March 9, 1976.
In line static mixers	Installed Jan. 27, 1976.
Function	(a) To mix chlorine and chlorine dioxide solutions before adding to unbleached pulp stream and (b) to mix the blended chlorine and chlorine dioxide witn unbleached pulp.
Mixer (a)	
Diameter	102 mm (4 in.)
Length	1.07 m (42 in.)
No. of mixing elements	6
Construction	Carbon steel, saran-lined, titanium elements.
Maintenance costs	Nil
Connected hp	Zero
Energy consumption	Δ P of 4.8 kPa (0.7 psi) at 15.8 L/s (250 gpm)
Type	Static, in-line
Mixer (b)	
Diameter	305 mm (12 in.)
Length	3.7 m (12 ft)
No. of mixing elements	12
Construction	FRP
Maintenance costs	Nil
Connected hp	Zero
Energy requirement	Δ P of 55 kPa (8 psi) at 12.6 L/s (2000 gpm)
Type	Static, in-line

Retention Towers—

Because the pulp mill water system was revised and closed enough to build the bleach plant without using more ground water, the temperature of the unbleached pulp feed to the bleach plant increased to about 52°C (125°F). Therefore, it was necessary to provide D/C stage equipment with short retention time.

D/C Stage Towers (2 in series)—

Diameter	2.44 m (8 ft)
Height	6.10 m (20 ft)
Type	Vertical cylindrical
Construction	Steel, FRP lined
Type of agitator	Full length vertical shaft with 6 blades
Agitator	Steel FRP covered shaft, cast iron blades construction
Time of retention	4.5 minutes each
Fiber consistency	3.5 percent

D/C Washer—

As related in Appendix C (paragraph heading "Twin roll press") the D/C stage of the oxygen bleach plant was operated from December 1972 to July 1975 without a D/C washer. Because of problems described in the referenced section the need for a D/C washer became clear.

D/C Washer Mound Dimensions—

Diameter	2.9 meters (9 ft 6 in.)
Length	4.3 meters (14 ft)
Equivalent filtering	37.2 m^2 (400 ft^2) area
Type	IMPCO CV-3
Construction	317 SS vat inlet and discharge, titanium vat, Kynar wire, SS wire support, and FRP seal tank
Type of take off	Airlift
Discharge fiber consistency	12 percent

Twin Roll Press (TRP)—

Preceding the oxygen stage, the TRP functions to increase fiber, or stock, consistency from 3.5 to 4.5 percent to 32 to 36 percent. Therefore, water removed is more than 22 m^3/o.d. metric ton (5000+ gal/ton), or 5984 m^3/day (1.6 mgd) at 272 metric a.d. tons/day (300 a.d. tons/day). An end view of the TRP is shown in Figure A—1. This shows the cross section of the vat, the risers, the cleanout ports, the structural design of the press rolls and the relative thickness of the drilled roll cover.

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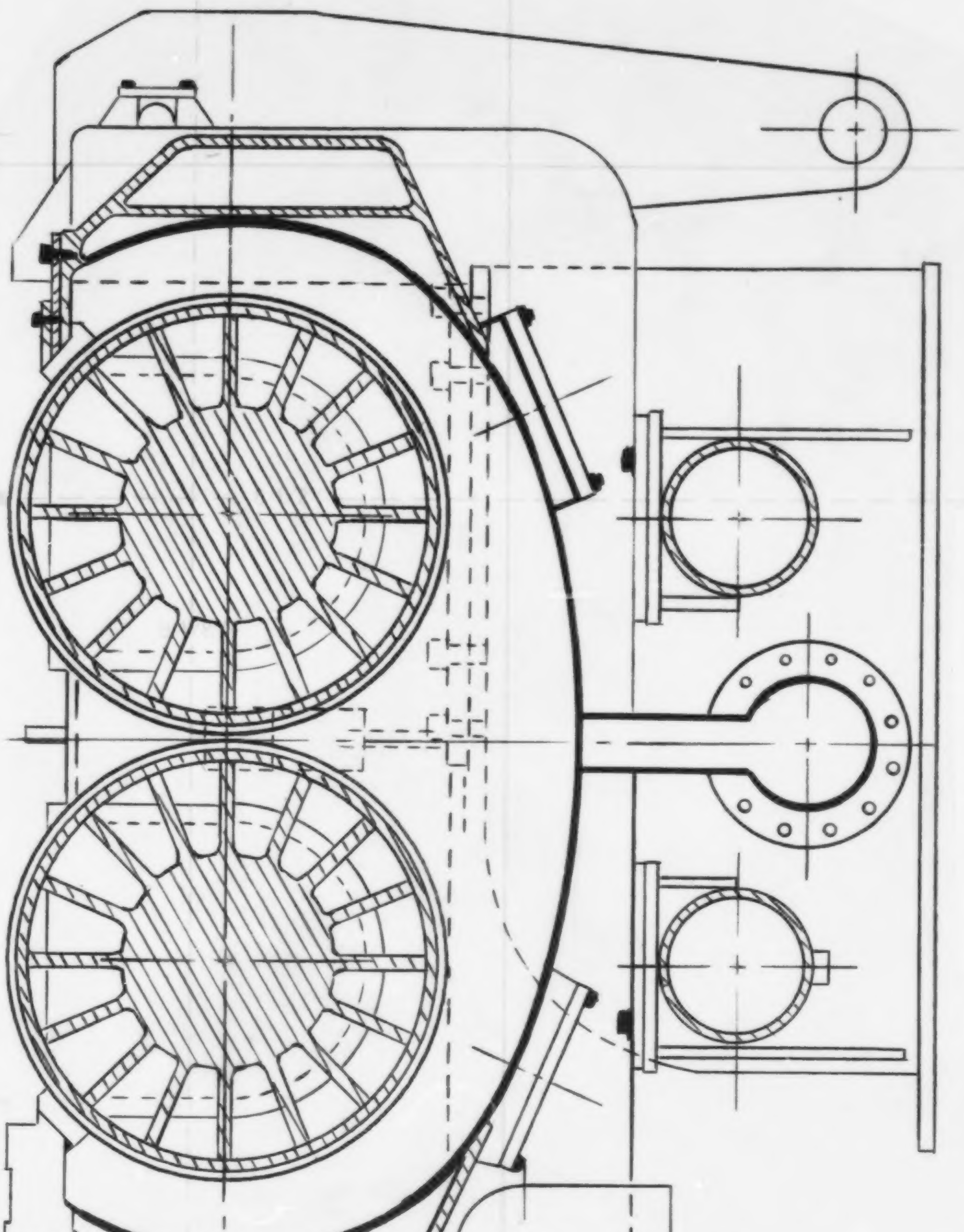

Figure A-1. End view of twin roll press.

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Vat pressure	82.7 to 103 kPa ga. (12 to 15 psig)
Alarm sounds at	138 kPa ga. (20 psig)
Feed pump stops at	165 kPa ga. (24 psig)
Rupture disc relieves at	207 kPa ga. (30 psig)
Feed pipe (riser) diameter	102 mm (4 in.) I.D.
Number of risers	4
Cleanout port diameter	0.44 meter (17.5 in.)
Number of cleanout ports	4 (2 each side)
Construction	317 SS drilled cover, 317 SS cover support, 317 SS roll structure, 317 SS vat, 317 SS breaker conveyor, and FRP pressate tank

Press Rolls—

Outer diameter	0.90 meter (35.43 in.)
Face length	2 m (78.74 in.)
Drilled cover thickness	4 mm (0.158 in.)
Drilled hole diameter	1 mm (0.04 in.)
Hole spacing, center to center	2 mm (0.08 in.)
Nip opening, minimum	0.4 mm (0.015 in.)
Nip opening, maximum	15.9 mm (0.625 in.)
Roll positions	One roll stationary, one roll moveable
Nip pressure	0 to 444 kN/m (0 to 2500 pli)
	Set by operator
	Controlled by automatic opening and closing of nip

Rotational Speed

Drive	Initial, 1750 rpm, 93.3 kW (125 hp)
	Present, 2200 rpm, 112 kW (150 hp) variable speed capability
Rolls	10 rpm max (normal operating condition, except 7 rpm for low production rate

Materials of Construction

Roll cover	317 ELC stainless steel
Roll shell	317 ELC stainless steel

Screw (or Transfer) Conveyor—

This conveyor, variously named in the records as screw, transfer or breaker conveyor, is located physically above the TRP.

Function	To pick up and convey 38 to 40 percent consistency pulp from the TRP to the steam mixer ahead of the screw conveyor which feeds to the O_2-stage reactor (normally conveys to steam mixer at 32 to 34 percent pulp consistency).
Dimensions	406 mm dia by 406 mm pitch by 6210 long mm (16 in. dia by 16 in. pitch by 20 ft - 4½ in. long)
Design capacity	1.42 m^3/s (3000 cfh) of 38 percent consistency pulp at 59 rpm
Construction (screw and trough)	317 stainless steel

Oxygen Stage

The second stage system receives pulp from the TRP at 32 to 34 percent consistency, heats it up and mixes it with caustic, compresses it, and feeds it as a seal plug to a device that breaks, or fluffs, the plug into small pulp bundles to provide accesibility to oxygen in the reactor. The reactor retains the alkaline pulp (downward movement) under presure a controllable time, discharges it into a dilution section at the bottom then blows it to an agitated tank. From the latter tank the pulp goes to the vacuum washer.

Single Shaft Steam Mixer—

Function	Steam heats pulp at 34 percent consistency to 87.8°C (190°F) and mixes it simultaneously with 29.3 kg/metric a.d. ton (65 lb/a.d. ton) of 7 percent caustic.
Length	2.67 meters (8 ft 9 in.)
Diameter	0.8 meters (2 ft 8 in.)
Construction	Horizontal shaft, 6 to 8 helical blades, 2 steam inlets per side, 6 chemical inlets from single maifold; 317 stainless steel
Type	Serial No. G-809

Screw Feeder—

Function	Conveys high density pulp from steam mixer to fluffer at top of reactor. Forms pulp plug to act as pressure seal for reactor.
Length	2540 mm (100 in.)
Diameter	Tapered from 591 mm (23.27 in.) at feed end to 381 mm (15 in.) 1397 mm (55 in.) from feed end; enlarges from there to 425 mm (16.73 in.) at end discharging to fluffer.

Type	IMPCO—Sunds, Serial No. J-143
Construction	317 stainless steel
Fluffer— Function	Breaks up pulp plug into small fragments to decrease diffusion path for gaseous oxygen to contact all pulp uniformly.
Rotational speed	990 rpm
Connected horsepower	37.3 kW (50 hp), 1800 rpm
Construction	Vertical tapered shaft and helical vanes with serrated edges; 316 stainless steel
Reactor— Function	Provides contact for selected time between fluffed pulp and oxygen gas at controlled pressure and temperature.
Dimensions	
Top section	Diameter—3.0 meters (9 ft 10 in.)
	Height—7.8 meters (25 ft 7 in.)
Bottom section	Diameter—3.5 meters (11 ft 6 in.)
	Height—4.3 meters (14 ft. 2 in.)
Construction	Body—steel
	Lining—Carpenter 20
Operating pressure	Initial—620.5 kPA (90 psig)
	Present—414 kPa (60 psig)
Safety Systems— Function	To provide protection against excessive temperature and pressure rise, fires and explosion in oxygen reactor.

System Components

A. A 2.5 cm (1 in.) hand operated vent valve to drop temperature and pressure

B. A 10 cm (4 in.) automatic pressure relief valve set at 965 kPa (140 psig)

C. A 30.5 cm (12 in.) rupture disc designed to relieve at 999 kPa (145 psig)

D. A 30.5 cm (12 in.) rupture disc designed to relieve at 1,034 kPa (150 psig)

E. A high pressure water deluge system that trips when upper or lower pulp bed level temperature reaches 143.3° C (290° F)

Oxygen Injection Points—

Function	Provides high-purity (99.5 percent or greater) oxygen gas to pulp in reactor.
Initial location	Four to 10 cm (4 in.) equally-spaced ports at periphery of lower end of top section, just above discharge table.
Present location	Four to 10 cm (4 in.) equally-spaced injection points near the top of the reactor to provide better contact with incoming fluffed pulp.

Discharge Devices—

Function	To move high-consistency oxygen bleached pulp to dilution section at lower end of bottom section of reactor
Location in reactor	Initial—see Figure A-2 Present—see Figure A-3
Construction	Initial—316 SS Present—317 L SS

Drives for Discharge Devices—

Function	Provides power and variable rotational system to move pulp discharge devices in oxygen reactor.
Type	(Initial) Hagglund 8385 hydraulic motor, 7,584 kPa (1100 psig) operating pressure: oil temperature 48.9 to 60° C (120 to 140° F); 20.7 to 27.6 kPa (3 to 4 psi) Δ P maximum across oil cooler; 0.19 to .25 L/s; (3 to 4 gpm) cooling water at 26.7° C (80° F) maximum and 861 kPa (125 psig) maximum with minimum velocity of 310 mm/s (1 ft/s). Present mechanical drive with helical gear ratio 12 to 1 and worm gear ratio 35.5 to 1.
Drive motors	(Initial) electric motor, TEFC, 37.3 kW (50 hp), 1800 gpm, frame 365 US, 3/60/550 (Present) D.C. motor, 75 kW (100 hp), TEFC, frame 502T, 1750 rpm, 480 volts with tach-generator mounted; solid stage power rectifier with seventeen controls and instruments for speed adjustment and indication; isolation transformer and operator control station.

Oxygen Blow Tank—

Function	Receives diluted, O_2-stage bleached pulp from reactor and serves as feed tank to oxygen stage washer
Dimensions	Diameter—4.9 m (16 ft) Height—5.5 m (18 ft)
Construction	Body—steel Lining—Flake Ceilcote 103 Agitator—316 SS

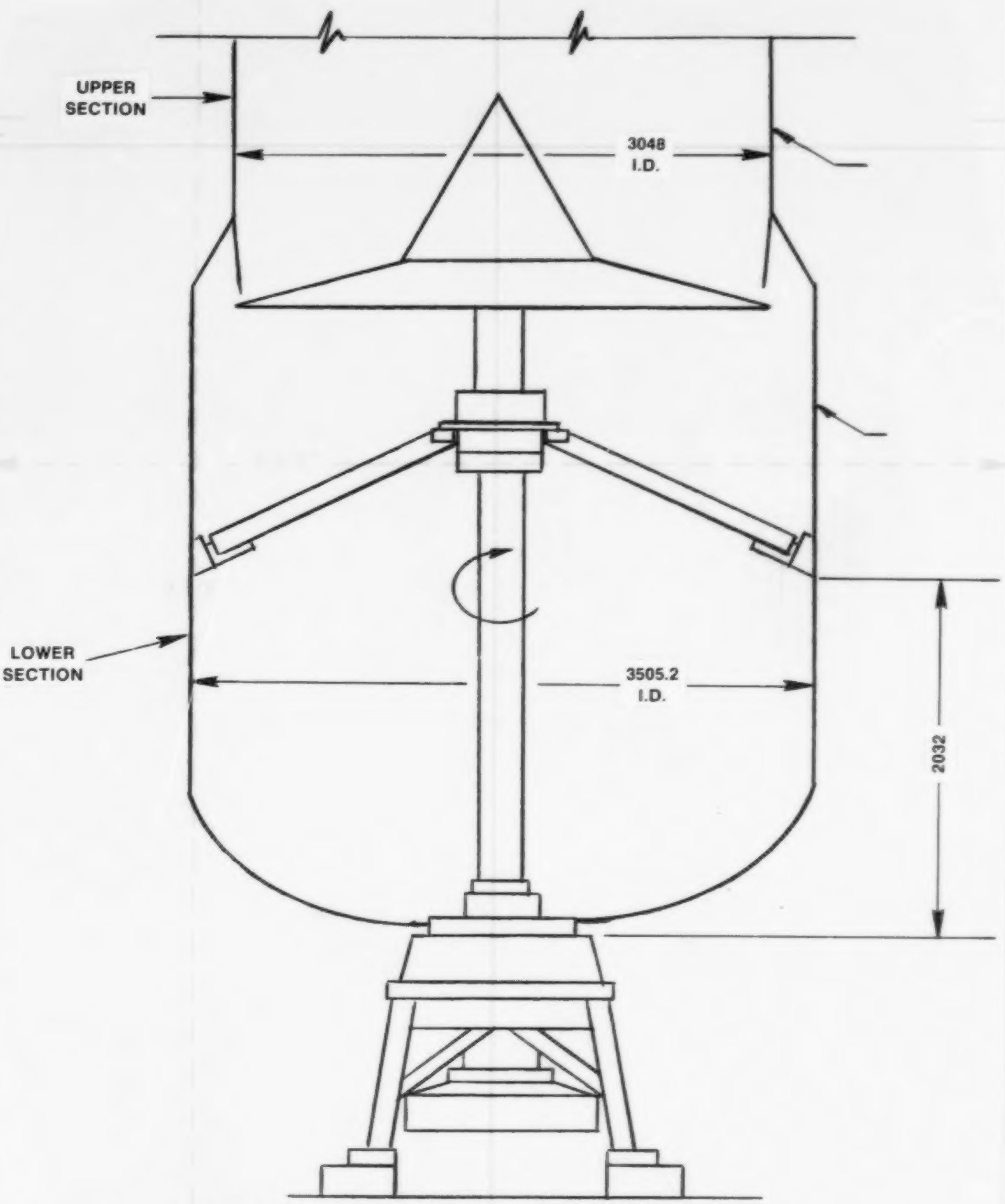


Figure A—2. **Initial discharge device in oxygen reactor. (Dimensions given in millimeters.)**

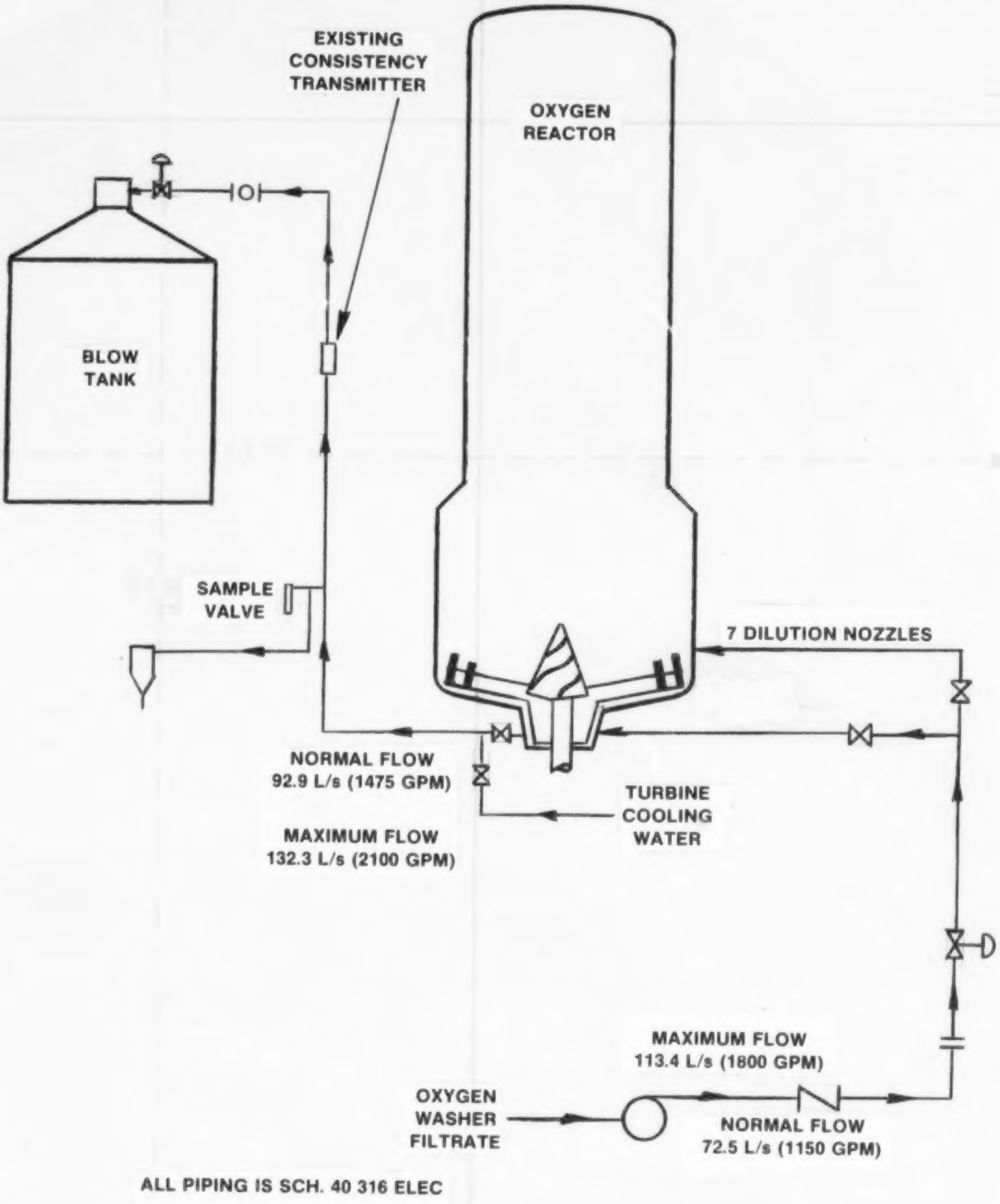


Figure A—3. Wet bottom discharger in oxygen reactor.

Oxygen Washer System—

Function	Washes O-stage bleached pulp, collects and recycles filtrate, thickens pulp ahead of D-stage
Washer mould dimensions	Diameter—2.9 m (9 ft 6 in.)
	Length—4.3 m (14 ft)
Type	IMPCO CV 3
Construction	Wire—Polyester
	Wire support—316 SS
	Vat—316 SS
	Seal tank—FRP

Double Shaft Steam Mixer—

Function	Raises temperature of oxygen stage bleached pulp ahead of D-stage and mixes any chemical used for pH control.
Dimensions	Horizontal width—1.6 m (5 ft 2 in.)
	Vertical width—1.0 m (3 ft 3 in.)
	Length—1.7 m (5 ft 7 in.)
Type	IMPCO Series 300-10, Serial No. H-817
Steam nozzles	2 (1 each side)
Construction	304 SS

D-Stage

The third stage system pumps the hot O_2-stage pulp at 16 percent consistency, mixes it with ClO_2 solution, passes it upward in an updraft tube then downward into a variable depth tower, therefore, controllable time and washes the pulp on a vacuum washer before sending it to a tile high-density storage tower.

Thick Stock Pump—

Function	Pumps hot 16 percent consistency O_2-stage bleached pulp to the ClO_2 mixer and updraft tube.
Type	IMPCO Series 400
Connected horsepower	75 kW (100 hp), 1200 rpm motor, 100 rpm pump speed
Construction	316 SS

Chlorine Dioxide Mixer—

Function	Mixes about 1 percent ClO_2 (on O.D. pulp basis) with 10 to 12 percent consistency O_2-stage bleached pulp from thick stock pump.
Dimensions	Diameter—81.2 cm (32 in.) Length—148.3 cm (58-⅜ in.) Inlet diameter—40.6 cm (16 in.) Outlet diameter—40.6 cm (16 in.)
Type	Single shaft, IMPCO series 400, Serial No. G-482

Updraft Tube—

Function	Conveys O_2-stage bleached pulp, mixed with ClO_2 to top of downdraft tower.
Dimensions	Diameter—1.8 m (5 ft 11¼ in.) Height—27.5 m (90 ft)
Retention time	Dependent on production rate—at 10 percent consistency and 272 a.d. metric tons/day (300 a.d. tons/day), retention time is 41 minutes.
Construction	Initial—FRP Replacement—FRP Second replacement—titanium, Grade 2, 50 mm (3/16 in.) thick welded sections of rolled plate

Downdraft Tower—

Function	Retains partial ClO_2—bleached pulp for variable retention time
Dimensions	Diameter—5.4 m (18 ft 10 in.) Height—27.5 m (90 ft)
Retention time	Depends on production rate and level held in tower—at 10 percent consistency, 272 a.d. metric tons/day (300 a.d. tons/day) and 18.3 m (60 ft) pulp level in tower, retention time is 250 minutes.
Construction	Initial—FRP Replacement top—FRP Present and future—FRP

D-Washer—

Function	Washes D-stage bleached pulp, collects and recycles filtrate, thickens pulp ahead of tile storage tower.
Dimensions	Diameter—2.9 m (9 ft 6 in.)

	Length—4.3 m (14 ft)
Type	IMPCO CV 3
Construction	Wire—polyester
	Wire support—317 SS
	Vat and drain—317 SS (replacement of same material installed June, 1978)
Discharge consistency	12 percent

High Density Pulp Storage

Function	Stores 181 metric tons/day (200 tons/day) of fully bleached pulp at 12 percent consistency. Receives pulp from Warren thick stock pump following the "D" washer; provides storage for wet end of pulp dryer.
Dimensions	See Figure A-4
Construction	Semblock tile exterior; Semblock plate and radial tile interior; concrete filling between inner and outer tile shells, interior portland cement joints. Cover of tower is Atlac 711-05A, high heat and chemical resistant FRP.

AIR SEPARATION PLANT

The term "air separation plant" is commonly used to designate plants designed to produce oxygen, nitrogen, and/or rare gases from air. At Chesapeake, the air separation plant is usually referred to as the on-site oxygen plant primarily because the oxygen it produces is the main product used by Chesapeake. There is some high-purity, gaseous nitrogen produced but, presently, this is used for cooling in the plant and then vented to the atmosphere.

The oxygen plant was designed to produce 90.7 metric tons/day (100 tons/day) of oxygen. Of the total, 81.6 metric tons/day (90 tons/day) was gaseous and 9.1 metric tons/day (10 tons/day) was liquid at a purity greater than 99.5 percent. Gas product pressure was 172 kPa (25 psig). Liquid product flowed to an insulated storage tank, described below, which floated on line to provide oxygen gas steam vaporized from liquid as required. Excess liquid oxygen was, and still is, sold by long-term contract to an oxygen supplier.

Tests made to establish the actual oxygen plant capacity and mechanical or other problems are described in Appendix C.

General information on the oxygen plant is listed below:

Plant designation	Model 0-10,000
Defrost time	24 hours
Warm start-up time	24 hours
Cold start-up time (short shutdown)	1 to 2 hours
Continuous operating time (without defrost)	12 months

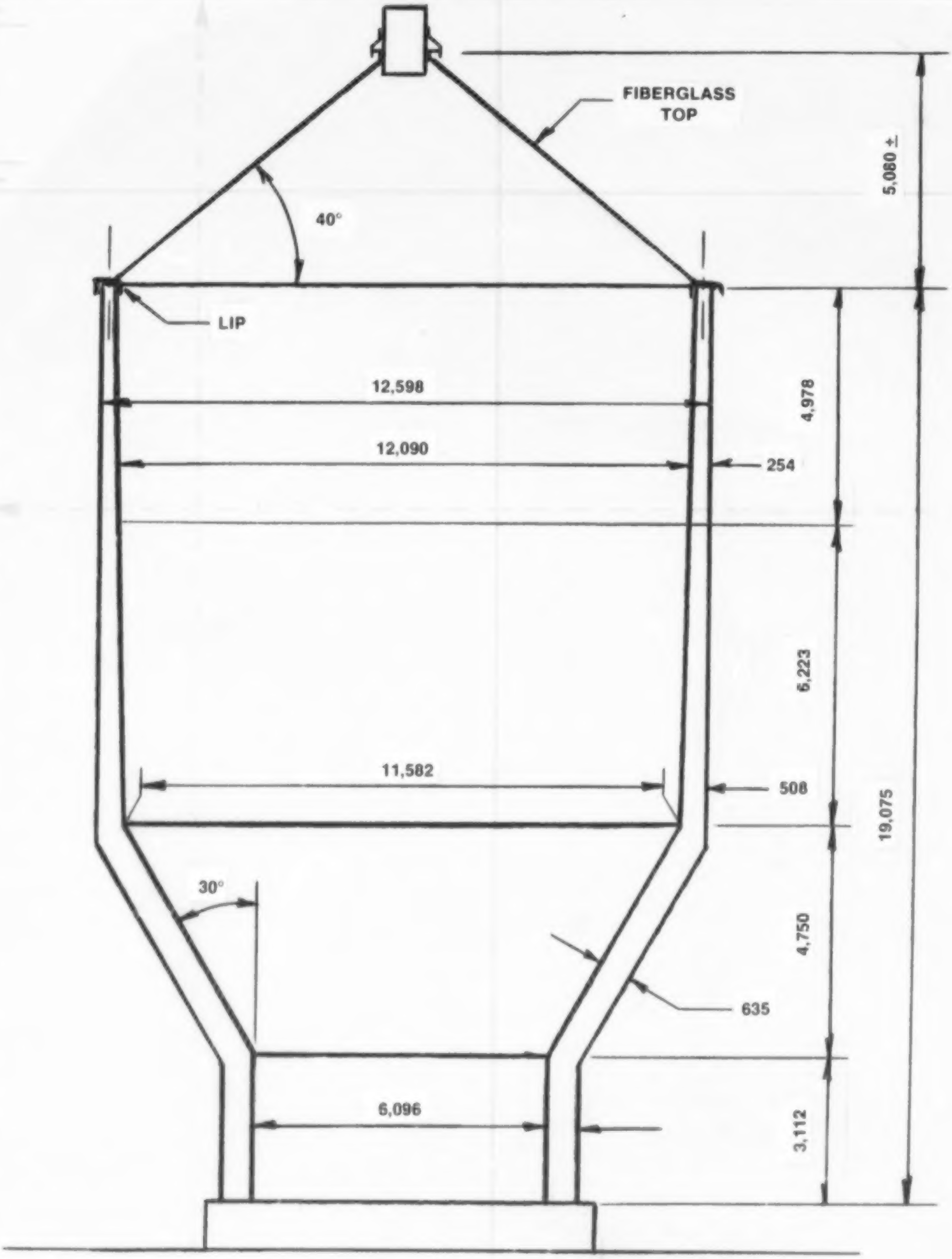


Figure A—4. High density bleached pulp storage chest. (Dimensions given in millimeters.)

Recommended defrost time	12 months

Following is a description of the major functional equipment in the oxygen plant system. This equipment is shown on the schematic flow diagram in Figure A-5.

Air Compressor

Function	Compresses air by four-stage compression with interstage cooling.
Type	Joy Turbo-Air 4MSG-6 Centrifugal Serial No. BF-5774

Compressor Rating

Gas	air
Relative humidity	58 percent
Inlet Pressure	99.2 kPa (14.4 psig)
Inlet volume capacity	313 m^3/min (11,049 cfm)
Inlet temperature	1st stage—32.2° C (90° F)
Inlet temperature	2nd stage—40.0° C (109° F)
Inlet temperature	3rd stage—42.8° C (109° F)
Inlet temperature	4th stage—46.1° C (115° F)
Discharge pressure	1120 kPa abs (162.5 psia)
Cooling water temperature	29.4° C (85° F)
Input speed	1785 rpm

Compressor Drive

Connected power	1865 kW (2500 hp)
Motor type	Squirrel cage induction, 2300 volt, 3 phase, 60 Hz, 1.15 service factor, 1800 rpm

Interstage Coolers—

Function	Cools compressed air after first, second and third stage of compression by heat exchange with cooled water from cooling tower.
Type	Tube and shell, exact dimensions not included in compressor drawings and specifications.
Shell side air flow	353 kg/min (773 lb/min) dry
Tube side water flow	16.5 L/s (262 gpm)
Tube side water temperature	Inlet—29.4° C (85° F) Outlet—37.8° C (100° F)

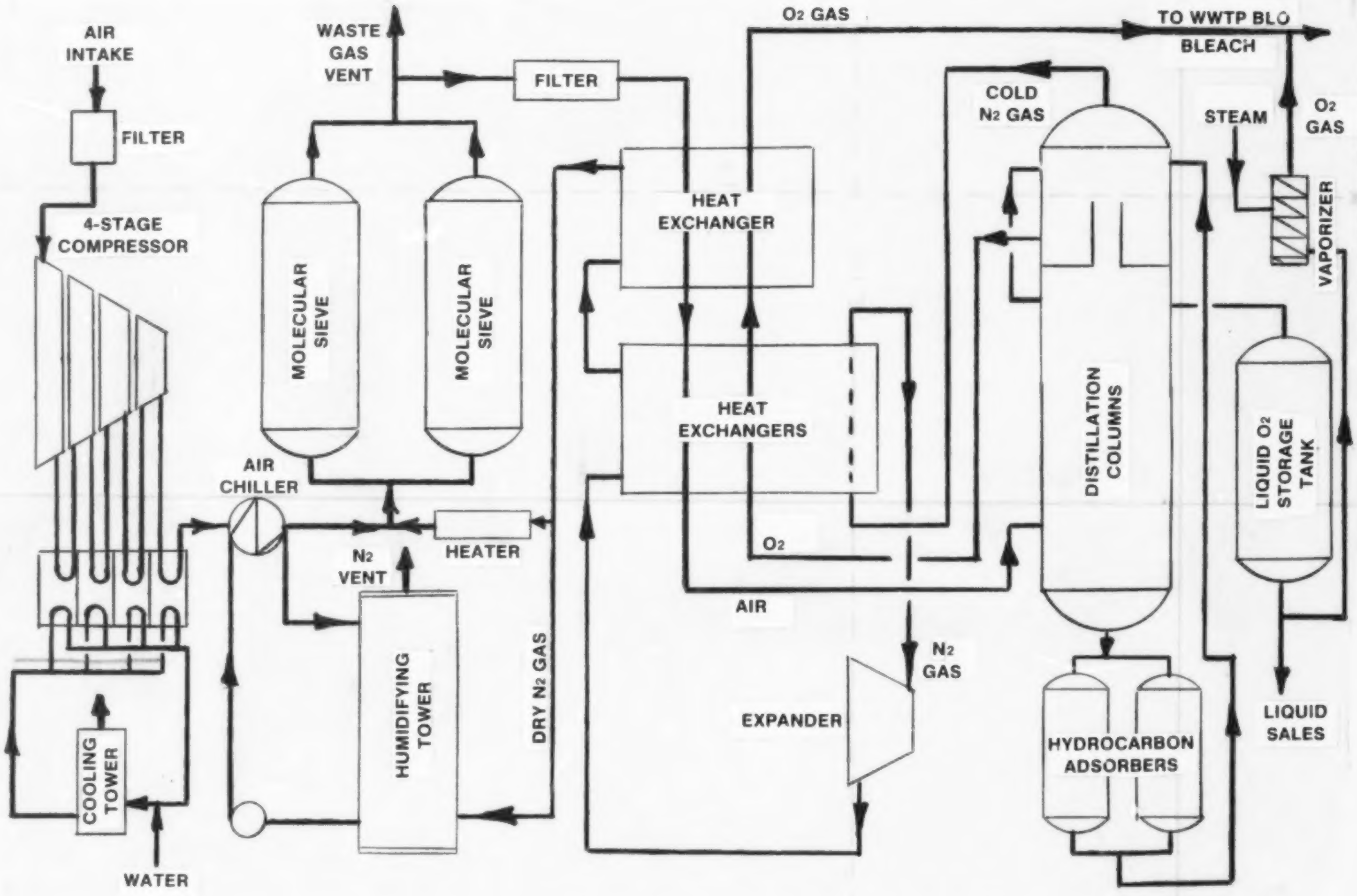


Figure A—5. Schematic flowsheet of oxygen plant.

Tube side pressure drop	76 kPa (11 psi)
Number of coolers	One each after first, second and third stage of compression.
After Cooler—	
Function	Cools air after fourth stage of compression by heat exchange with cooled water from cooling tower.
Type	OP, Size 16-168 shell and tube, removable tube bundle
Heat transfer surface	94 m^2 (1013 ft^2) per shell
Shell side air volume	302.4 m^3/min (10,800 cfm) at pressure and temperature at compressor inlet
Shell side temperature	Design—121.1° C (250° F)
	In—94.4° C (202° F)
	Out—38.8° C (100° F)
Shell side pressure	Design—1206 kPa (175 psia)
	Test—1826 kPa (265 psia)
	Operating—1120 kPa (162.5 psia)
Shell side pressure drop	10.3 kPa (1.5 psi)
Shell construction	Steel shell
	Cast iron heads
Shell dimensions	Diameter—40.6 cm (16 in.)
	Length—4.58 m (15 ft.) including return end
Type baffles	Segmental
Baffles constructed of	Steel
Tube side water volume	9.5 L/s (150 gpm)
Water Temperature	In—29.4° C (85° F)
	Out—37.8° C (100° F)
Tube pressure	Design—1034 kPa (150 psia)
	Test—2067 kPa (300 psia)
Tube pressure drop	20.7 kPa (3 psi)

Tube dimensions	O.D. — 22.9 mm (¾ in.) Length—4270 mm (14 ft) BWG—18
Tube construction	Inh. Adm. LoFin

Cooling Tower

Function	Cools water recirculating to interstage coolers aftercooler and oil cooler at air compressor.
Type	Single cell, double cross-flow
Model	374-101, Series 10
Constructed of	Treated wood, corrugated asbestos cement casing and louvers
Dimensions	Width—6.25 m (20 ft 6 in.) Length—7.06 m (23 ft 2 in.) Height—3.91 m (12 ft 11 in.) Filling—1525 m^2 (16,410 ft^2) Splash surface—481 m^2 (5200 ft^2) Effective cooling volume—40.8 m^3 (1440 ft^3)

Operating Parameters

Water flow	69.3 L/s (1100 gpm)
Inlet temperature	37.8°C (100°F)
Outlet temperature	29.4°C (85°F)
Design wet bulb temperature	25.6°C (78°F)
Drift loss	0.2 percent
Evaporative loss	1.5 percent
Tower loading	14.9 m^3/m^2h (6.1 gpm/ft^2)
Fan diameter	3.05 m (10 ft)
Fan blades	9
Fan connected power	26.9 kW (36 hp)

Dryer System—CO_2 and H_2O Removal

Function	Two molecular sieve beds, alternately "on-stream" or being regenerated with heated nitrogen (N_2); gas removes carbon dioxide (CO_2) and water (H_2O) from the process air stream.
Vessel dimensions	Diameter—2.02 m (6.63 ft.)
	Height—2.28 m (10 ft 9 in.)
Regeneration System	Gas heaters—two electric at 150 kW each
Bed load	Approximately 4950 kg, (11,000 lb)/bed leveled to + 25 mm (1 in.) on top of bed.
Operating pressure	Lowest expected is 1,034 kPa (150 psig)
Outlet temperature	260° C (500° F)

Dust Filter

Function	Removes dust carry-over from molecular sieve vessels
Total filtering area	6.04 m^2 (65 ft^2)
Filter media	proprietary information
Dimensions	Diameter—406 mm (16 in.)
	Height—2.26 m (89 in.)
Test pressure	1550 kPa (225 psig)
Maximum operating pressure	1,034 kPa (150 psig) at 260° C (500° F)
Construction	Vertical carbon steel shell 4.8 mm (3/16 in.) thick

Hydrocarbon Adsorbers (Two Vessels)

Function	Removes principally acetylene from cold air stream
Dimensions	Diameter—508 mm (20 in.)
	Height—1.98 m (6 ft 6 in.)
Test pressure	1757 kPa (255 psig)
Maximum operating pressure	1137 kPa (165 psig) at 204.4° C (400° F)
Minimum temperature	-196° C (-320° F)
Construction	Vertical stainless steel shell 3.2 mm (⅛ in.) thick
Adsorbent	Silica gel

Distillation Columns

Function	Separates liquid air into gaseous and liquid oxygen (O_2) and gaseous nitrogen (N_2)

Lower Column—

Dimensions	Diameter—1.53 m (5 ft)
	Height—4.27 m (14 ft)
Test presure	2000 kPa (290 psig)
Maximum operating pressure	1206 kPa (175 psig) at maximum temperature of 37.8°C (100°F) and minimum temperature of -196°C (-320°F)
Construction	Vertical stainless steel shell 9.5 mm (⅜ in.) thick
Insulation and internals	Proprietary information

Upper Column—

Dimensions	Diameter—1.83 m (6 ft)
	Height—10.1 m (33 ft)
Test pressure	552 kPa (80 psig)
Maximum operating pressure	345 kPa (50 psig) at maximum temperature of 37.8°C (100°F) and minimum temperature of -196°C (-320°F)
Construction	Vertical aluminum shell 9.5 mm (⅜ in.) thick
Insulation and internals	Proprietary information

Liquid Oxygen (LOX) Storage and Vaporizer System

Function	Receives LOX from oxygen plant as backup supply for normal gaseous oxygen produced. Steam vaporizes LOX into gaseous oxygen supply line to WWTP, BLO and oxygen bleach plant as required.

LOX Storage Tank—

Dimensions	Diameter—3.05 m (10 ft) outside
	Length—horizontal cylindrical elevated, 18.3 m (62 ft) outside length
Capacity	105.9 m^3 (28,000 gallons) or 120.8 metric tons (133.2 tons)
Operating pressure	Design—345 kPa (50 psig)
	Actual—172 kPa (25 psig)
Operating temperature	-212°C (-350°F)
Construction	Carbon steel outer shell, stainless steel inner shell, stainless steel and brass flanges and pipelines

Insulation	Perlite and vacuum

LOX Vaporizer—

Type	Tube and shell
Tube pressure	689 kPa (100 psig) at 170°C (338°F)
Snell pressure	689 kPa (100 psig) at -189°C (-300°F)
Steam pressure	414 kPa (60 psig)
Steam consumed	0.2 kg/s (1570 lb/hr)
LOX evaporated	2830 m^3/h (100,000 SCFH) or 90.7 metric tons (100 tons/day)
Construction	Stainless steel

Oxygen Distribution System

Oxygen produced by the cryogenic oxygen plant is stored, compressed and transported in a number of ways at Chesapeake. Storage at the oxygen plant itself has just been described above. That storage and other parts of the oxygen distribution system at Chesapeake are shown schematically in Figure A-6 and described in the following sub-paragraphs. The dotted line system shown to provide oxygen to the lime kiln has been installed but not operated as of October 31, 1978. Therefore, the lime kiln system will not be described in Appendix A or elsewhere in this report. The dotted line system for strong BLO has not been installed and, likewise, will not be reported.

Oxygen compressor—Bleach Plant—

Function	Compresses dry oxygen gas from 172 kPa (25 psia) to 1,034 kPa (150 psia) for injection into the oxygen reactor, or O-stage, in the bleach plant.
Type	No. 2E130-2A (Drawing No. 6-4050.001) (Equipment No. 26-65-01)

Description Data

Number of stages—2

Number of pistons—2 (single acting)

Piston stroke—130 mm (5.12 in.)

First stage piston diameter—275 mm (10.82 in.)

Second stage piston diameter—150 mm (5.9 in.)

Gas to be handled—dry oxygen

Suction volume—2.12 m^3/m (75 acfm)

First stage suction pressure—172 kPa (25 psig)

First stage suction temperature—43.3°C (110°F)

First stage discharge pressure—579 kPa (84 psia)

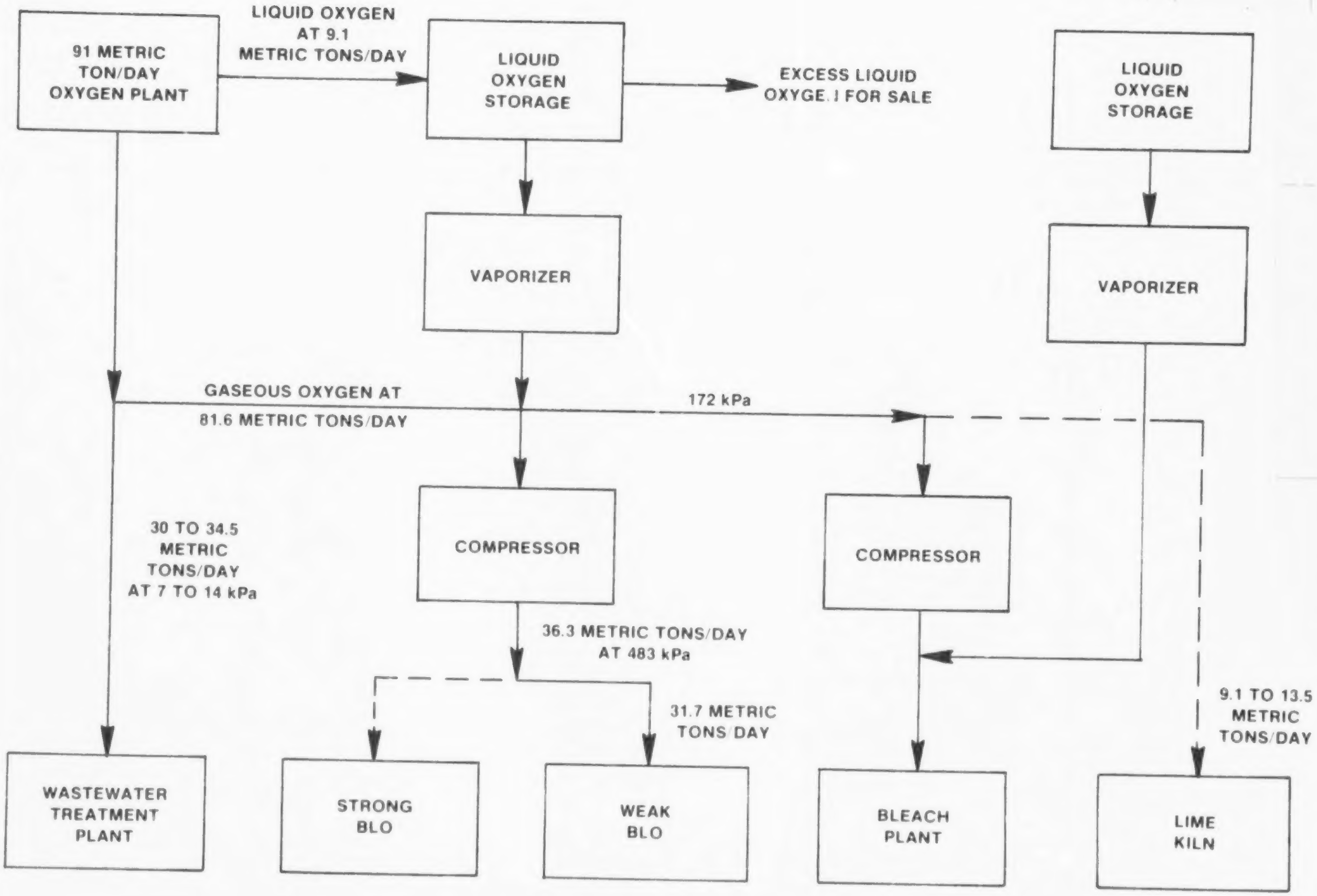

91 METRIC TON/DAY OXYGEN PLANT
LIQUID OXYGEN AT 9.1 METRIC TONS/DAY
LIQUID OXYGEN STORAGE
EXCESS LIQUID OXYGE.I FOR SALE
LIQUID OXYGEN STORAGE
VAPORIZER
VAPORIZER
GASEOUS OXYGEN AT
81.6 METRIC TONS/DAY
172 kPa
COMPRESSOR
COMPRESSOR
30 TO 34.5 METRIC TONS/DAY AT 7 TO 14 kPa
36.3 METRIC TONS/DAY AT 483 kPa
31.7 METRIC TONS/DAY
9.1 TO 13.5 METRIC TONS/DAY
WASTEWATER TREATMENT PLANT
STRONG BLO
WEAK BLO
BLEACH PLANT
LIME KILN

Figure A—6 Oxygen distribution system.

	First stage discharge temperature—174°C (345°F)
	Second stage suction temperature—35°C (95°F)
	Second stage discharge pressure—1206 kPa (175 psia)
	Second stage discharge temperature—107.2°C (225°F)
	Temperature after aftercooler—35°C (95°F)
	Cooling water inlet temperature—26.7°C (80°F)
	Cooling water temperature rise—8.3°C (15°F)
	Cooling water consumption—9.976 m^3/ (20 gpm)
	Compressor speed—400 rpm
	Net power input at flywheel—23.1 kW (31 hp)
	Motor rating—29.8 kW (40 hp)/service factor 1.15
	Motor speed—1800 rpm
	316 stainless steel

Oxygen Storage—Bleach Plant—

Function	To provide backup LOX storage at the bleach plant in case of compressor or main oxygen plant malfunction.
Type	Vertical cylindrical, TM 3000 to 3505
Dimensions	Diameter—2.16 m; 2.44 m (85 in. I.D.; 96 in. O.D.)
	Height—3.68 m; 4.86 m (145 in. inside; 191¼ in. outside)
Shell thickness	1.23 cm (0.485 in.)
Top and bottom	1.19 cm (0.468 in.)
Capacity	11.37 m^3 (3000 gallons)
Pressures	Test—2811 kPa (408 psi)
	Maximum allowable—1724 (250 psi)
	Operating—551 kPa (80 psi) at 65.6°C (150°F)
Operating Temperature	Minimum— -196°C (-320°F)
Construction	Shell—Nickel steel (T.S. 100,000) SA553
	Heads—Nickel steel (T.S. 100,000) SA353

Insulation	Double wall—vacuum powder

LOX Vaporizer—Bleach Plant—

Function	To vaporize LOX to provide gaseous oxygen to the O-stage of the bleach plant in case of oxygen compressor malfunction; also provides rapid-fill capability for the O-stage reactor during bleach plant startup.
Type	Starfin ambient air vaporizer—low presure, SV-4X4 arranged in two banks SV-4X20 and SV-4X8
Pressures	Test—6,895 kPa (1000 psia)
	Burst—27,580 kPa (4000 psia)
	Maximum operating—5,170 kPa (750 psia)
	Actual operating—1034 kPa (150 psia)
	Pressure drop—13.8 kPa (2 psi) for a single SV-4X4
Temperatures	Working range— -250 to 52°C (-425 to 125°F)
	Discharge gas—within -6.7°C (20°F) of ambient air temperature
Structural features	Tube spacing 178 mm (7 in.) center to center on corners of square.
	Single tube length—2.04 meters (80½ in.)
	Number fins per tube—8
Construction	Aluminum alloy 6063-T5

Oxygen Compressor—Black Liquor Oxidation (BLO)—

Function	To compress gaseous oxygen from main supply line at about 138 to 483 kPa (20 to 70 psig) as supply to weak black liquor oxidation system
Type	H6
Capacity	Nominal—14.5 m^3 (513 acfm) at 68.9 kPa (10 psig) or 10.3 m^3 (365 acfm) at 138 kPa (20 psig) inlet pressure discharging at maximum pressure of 414 kPA (60 psig)
	Tested—(28) 53.6 metric tons/day (59.1 tons/day) or 11.8 m^3 (417 acfm) at 138 kPa (20 psig) inlet pressure discharging at 483 kPa (70 psig)
	Drive—explosion-proof motor, 150 kW (200 hp) connected, 1750 rpm, direct connected
	Sealing water—volume 1.89 to 3.15 L/s (30 to 50 gpm)
	Pressure—414 kPa (60 psig)
	Temperature—48.9°C (120°F)

Construction	Cast iron housing; bronze rotor, steel water separator with bronze fitted ball float valve
Pressure change across compresor	345 kPa (50 psig)
Oxygen Pipeline—	
Function	To convey oxygen gas to bleach plant and BLO compressors from oxygen plant
Length	549 m (1800 ft)
Pipe diameter	152 mm (6 in.)
Oxygen gas velocity	15.9 m/s (52 ft/s) at 90.7 metric tons/day (100 tons/day) and 172 kPa (25 psig) (actually the pipeline will never convey more than 63.5 metric tons/day (70 tons/day) which represents a maximum velocity well below the 61 m/s (200 fps) maximum recommended
Construction	853 carbon steel

WASTEWATER TREATMENT SYSTEM

The wastewater treatment system installed by Chesapeake in 1973—74 represented not only Best Practical Technology for conventional pollutants; in Chesapeake's opinion it was and still is, the Best Available Technology for secondary treatment of industrial wastes waters. The overall system, shown schematically in Figures A-7, A-8, and A-9 collects numerous waste streams and occasional, accidental spills, then treats them by primary clarification, cooling, bio-degradation in a high-purity oxygen activated sludge reactor, secondary clarification, and discharge of the treated effluent to the receiving stream through a carefully-designed bottom diffuser. Strong acid or alkaline spills are diverted to strong waste tanks and fed back slowly to the main waste stream.

Excess sludge, which is a combination of primary and secondary clarifier underflows, is dewatered by vacuum filters and truck-hauled to landfill.

In-Plant Collection System (See Figure A-8)

Sewers in the mill were revised and grouped so as to feed into four sumps initially; a fifth was added in 1977—78. Most storm sewers were separated from process sewers to prevent large volume run off from passing through the WWTP; a few believed at the time to be insignificant are to be separated in the future.

Special collection and control sub-systems were made a part of the overall system to collect surge volumes from spills, from equipment cleaning and from storage chests which are infrequently emptied because of paper mill or pulp mill outages or planned shutdowns.

Individual Sumps (See Figures A-10, A-11, and A-12)

Each sump serves a particular section of the mill as indicated in the sections below. Dimensions and line drawings of the sumps are shown in Figures A-10, A-11, and A-12. Because of problems with sump pumps as described in Appendix C, characteristics of the sump pumps and motors are also described in the following sections.

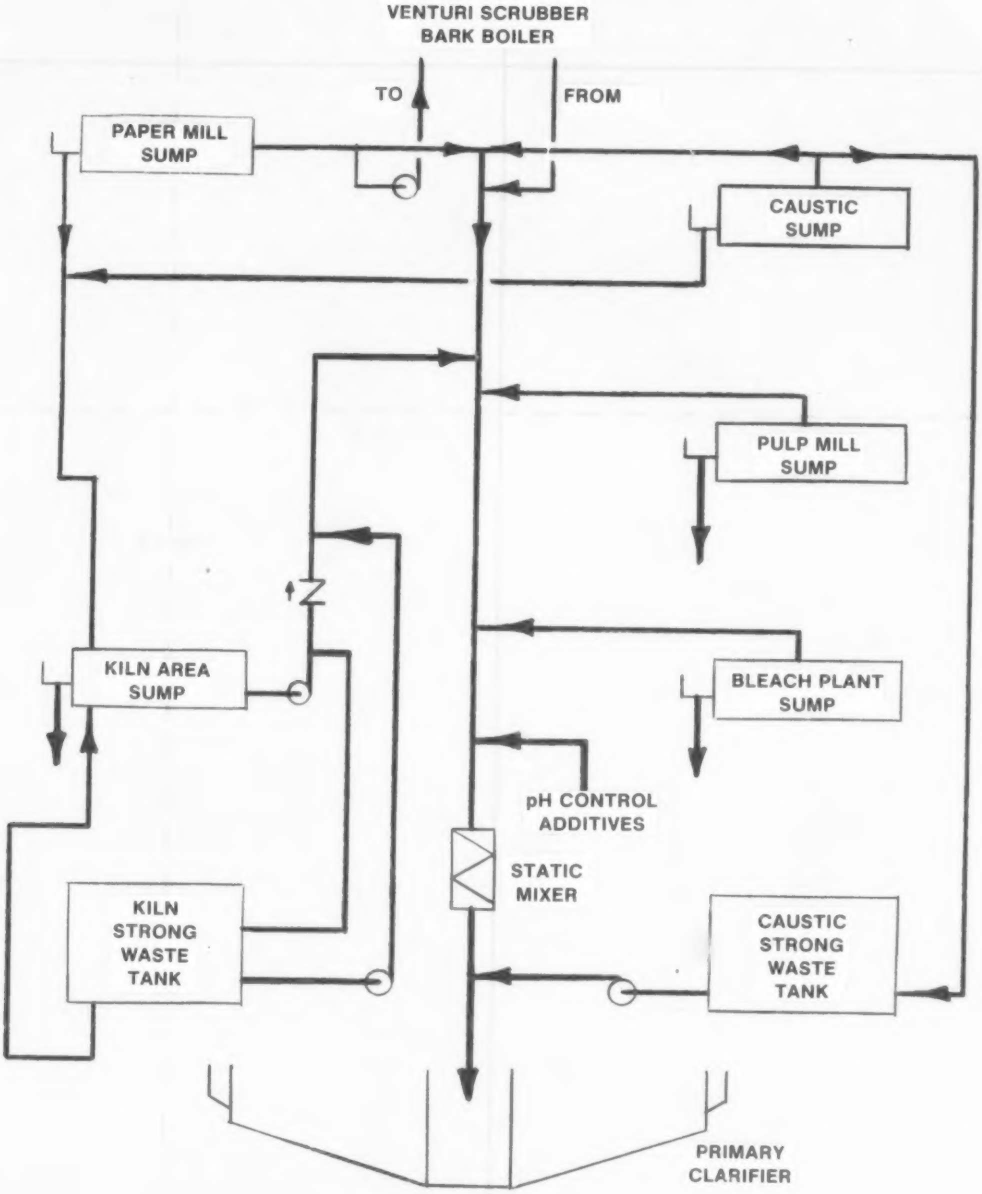


Figure A—7. Schematic flow diagram of wastewater collection system.

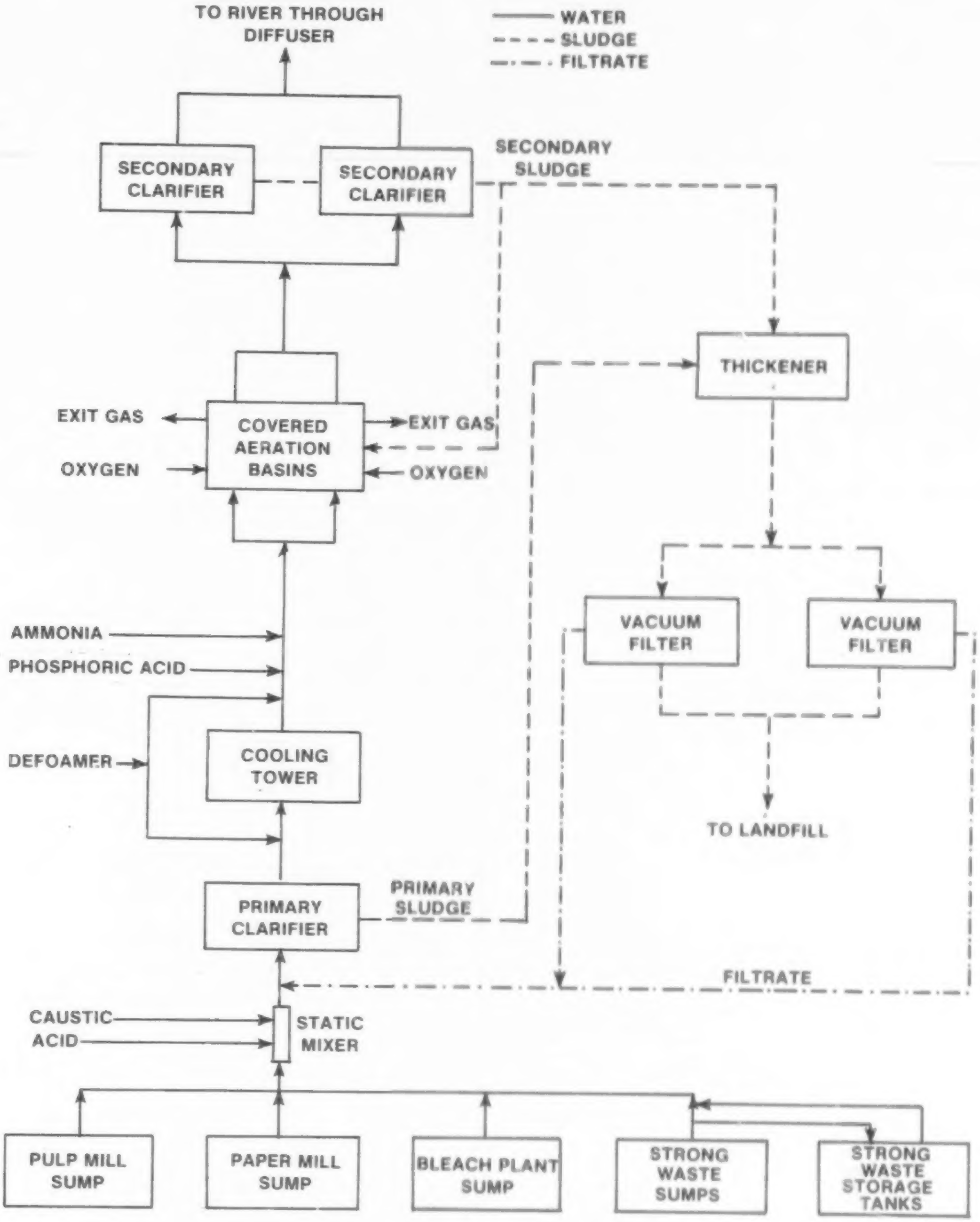
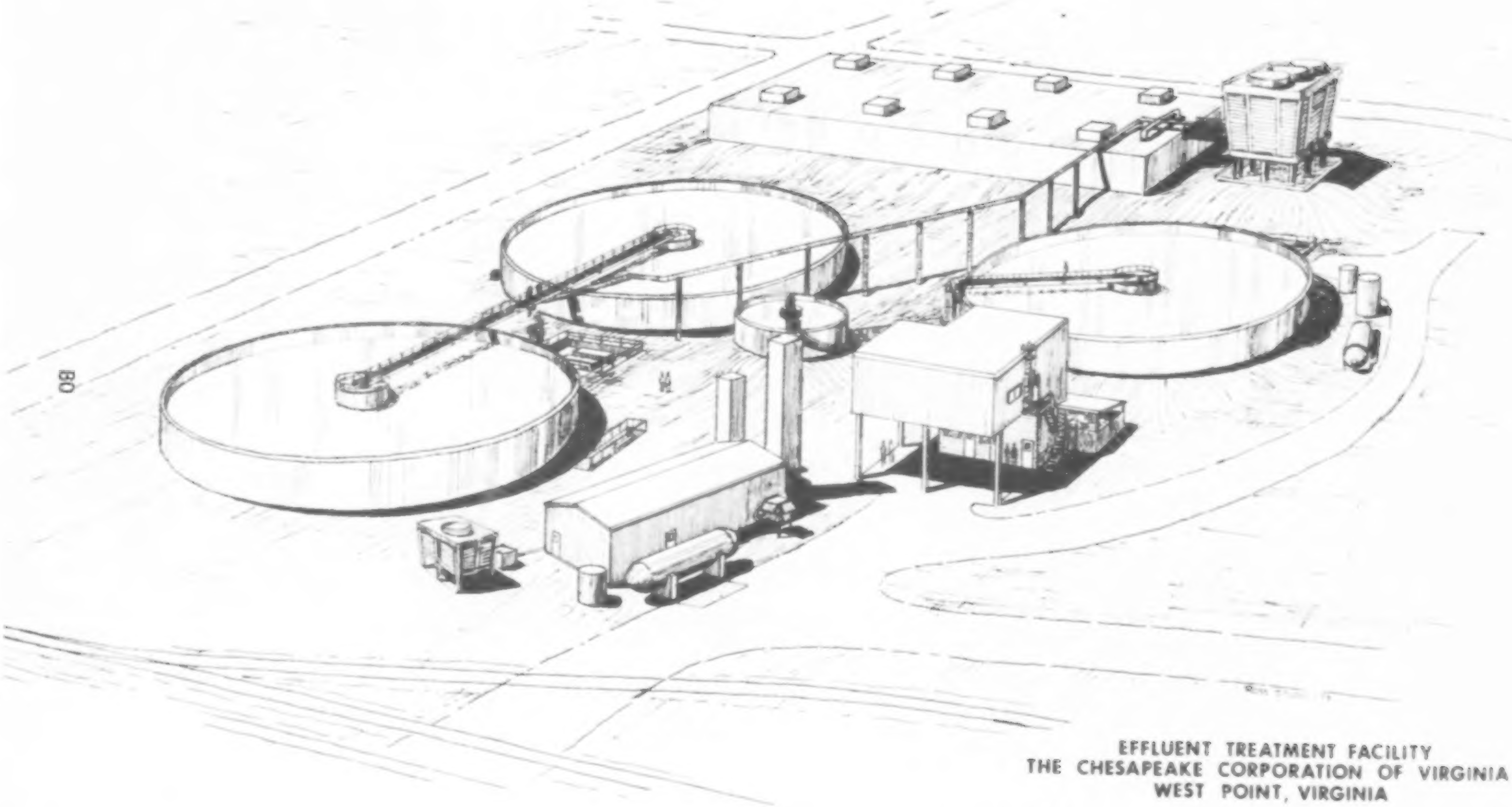


Figure A—8. Wastewater treatment plant flowsheet.



EFFLUENT TREATMENT FACILITY
THE CHESAPEAKE CORPORATION OF VIRGINIA
WEST POINT, VIRGINIA

Figure A—9. Perspective view of wastewater treatment plant.

Paper Mill Sump (See Figure A-10)—

Function	To receive wastewater from unbleached linerboard and paper machines, roll grinder room and hydro-pulper in the secondary fiber system; screens and pumps waste water to WWTP.
Sump dimensions	See Figure A-10
Construction	Reinforced concrete

Bar Screen

Type	Self-cleaning catenary, with trapezoidal bars, Peduro chain and angle iron cleaning bars
Drive system	0.56 kW (¾ hp) motor, 1800 rpm through L743B reducer to provide 7.5 rpm output
Overall width	0.92 m (3 ft)
Overall length	5.85 m (19 ft 2 in.)

Pumps and Motors

Motors (initial)	Two vertical motors, vertical hollow shaft, (VHS) totally enclosed fan cooled (TEFC) non-overloading with 1.15 service factor (SF); 3 phase, 60 Hz, 575 volts, 93.3 kW (125 hp), 1200 rpm
Pumps (initial)	Two vertical turbine pumps, Model No. 22HH-670 cast iron with 11 to 13 percent chrome impeller, trim and shaft; rated at 394 L/s (6250 gpm) at 18.3 m (60 ft) total developed head (TDH); would pass 25.4 mm (1 in.) solids.
Motor (replacement)	One of initial motors replaced with vertical solid shaft (VSS), high thrust motor with Class B insulation, 1.0 SF, TEFC, NEMA P base, AZV drip cover; 223.8 kW (300 hp) 1200 rpm, 2300 volts.
Pump (replacement)	One of initial pumps replaced with a 14 by 12 by 21 PWOV, CISS pump, with cutlass rubber shaft bearings for water lubrication; rated at 441 L/s (7000 gpm) at 35 meters (115 ft) TDH; would pass 70 mm (2-¾ in.) solids.

Pulp Mill sump (See Figure A-10)—

Function	To receive wastewater from areas including pulp washing, screen room, batch washer filtrate tanks, black water tank overflow, turpentine recovery system and tall oil sump; screens and pumps combined waste waters to WWTP.
Sump dimensions	See Figure A-10
Construction	Reinforced concrete
Bar screen	Identical to bar screen installed in and described above under the heading: Paper Mill Sump.

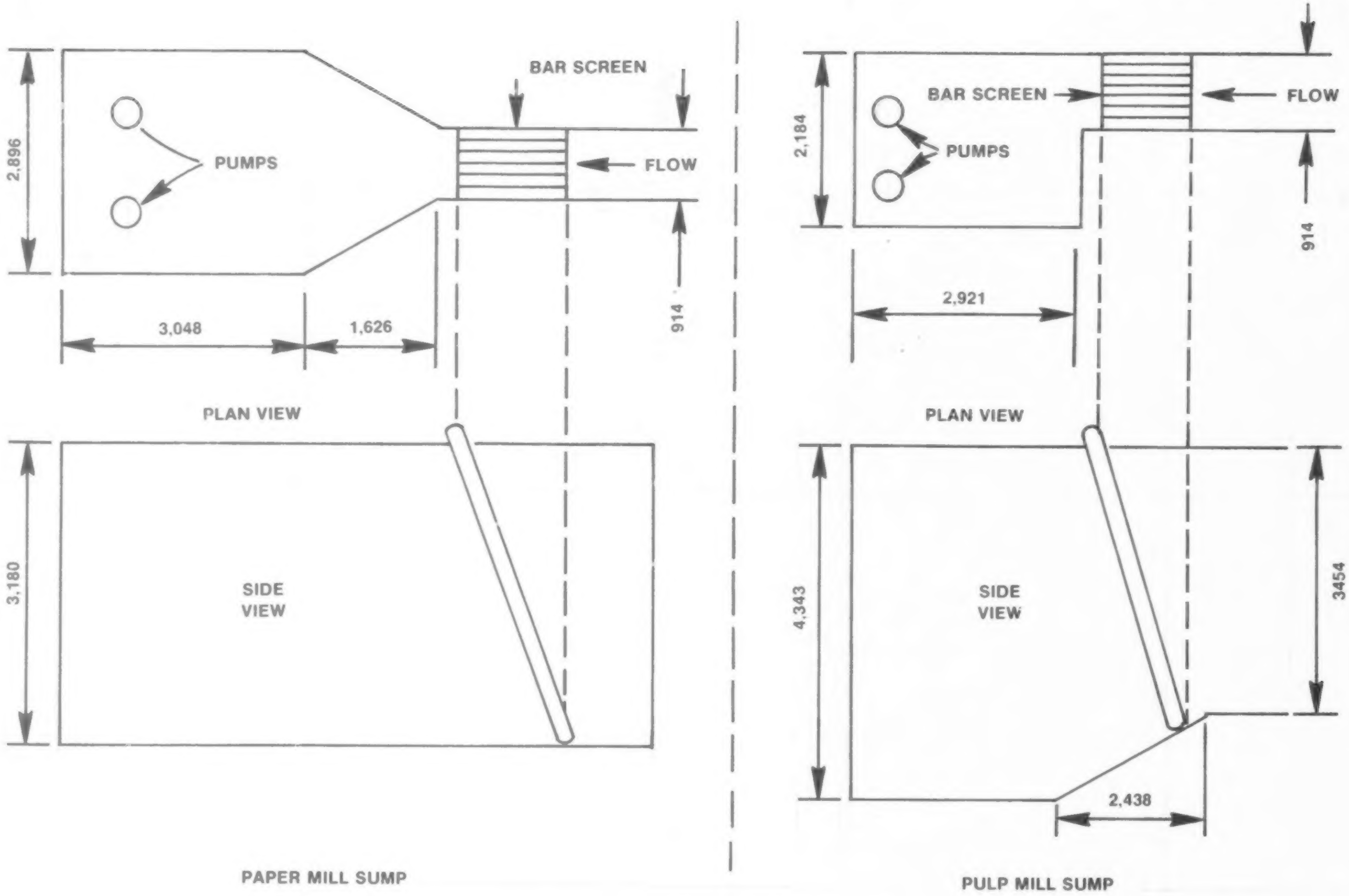


Figure A—10. Schematic view of paper mill sump and pulp mill sump. (Dimensions given in millimeters.)

Pumps and Motors

Motors (initial)	Two vertical motors, VHS, TEFC, non-overloading with 1.15 SF, 3/60/575 volts, 75 kW (100 hp), 1800 rpm
Pumps (initial)	Two vertical turbine pumps, Model No. 20H-430 cast iron with 11 to 13 percent chrome impeller, trim and shaft rated at 265 L/s (4200 gpm) at 22.3 m (75 ft) TDH, would pass 25.4 mm (1 in.) solids.
Motor (replacement)	One of initial motors replaced at pulp mill sump with one of the 93.3 kW (125 hp) 1200 rpm motors initially installed at paper mill sump.
Pump (replacement)	One of initial pumps replaced with a 12 by 10 by 17 SPWOV, maximum impeller, CISS pump to include pipe sections from pump to mounting plate; cutlass rubber shaft bearing for water lubrication; rated at 221 L/s (3500 gpm) at 27.5 m (90 ft) TDH; would pass 70 mm (2-¾ in.) solids.

Caustic Sump (See Figure A-11)—

Function	To receive waste water from Nos. 3 and 4 recovery boilers, the batch blow tank, causticizing and evaporator areas; screens and pumps wastewater to WWTP.
Sump dimensions	See Figure A-11
Construction	Reinforced concrete
Bar screen (initial)	Manually-cleaned bar screen 1.8 m (6 ft) wide with 13 mm (½ in.) openings between bars.

Bar Screen (Replacement)

Type	75° "MAX" mechanically-cleaned screen with rate of 9.5 mm by 51 mm (⅜ in. by 2 in.) steel bars with ½ inch openings mounted on 121 mm (4-¾ in.) thick cleaning rakes equally spaced on chains.
Drive system	0.75 kW (1 hp) motor, 1750 rpm through L743B reducer to provide 7.5 rpm output; 3/60/575 volts
Overall width	0.86 m (2 ft 10 in.)
Length, sprocket-sprocket	4.85 m (15 ft 11 in.)
Designed to operate	122.5 L/s (2.8 mgd) maximum flow (15 1.53 m (5 ft) water depth in sump) to 39.4 L/s (0.9 mgd) (at average 0.92 m (3 ft) water depth in sump); pH of 6.0 to 14.0; temperature of 37.8°C (100°F) to 65.6 (150°F); completely removable from sump.

Pumps and Motors

Motors (initial)	Two vertical motors, VHS, TEFC, non-overloading with 1.15 SF, 3/60/575 volts, 44.8 kW (60 hp) 1800 rpm
Pumps (initial)	Two vertical turbine pumps, Model No. 15-M-185 cast iron with 11 to 13 percent chrome impeller, trim and shaft, rated at 122.2 L/s (1940 gpm) at 27.5 m (90 ft) TDH; would pass 25.4 mm (1 in.) solids.

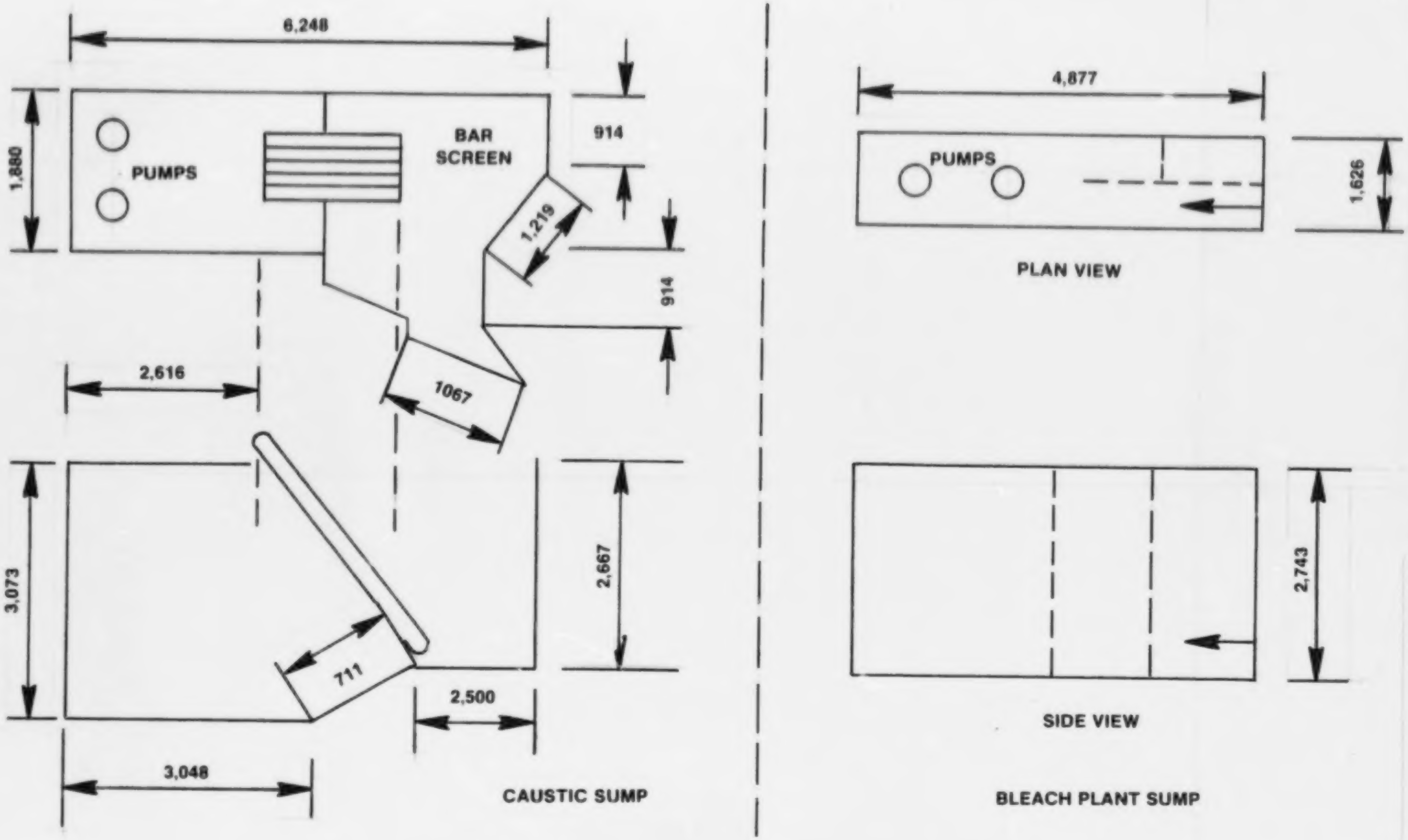


Figure A—11. Schematic view of caustic sump and bleach plant sump. (Dimensions given in millimeters.)

Motor (replacement)	One of initial motors replaced at caustic sump with one of the 75 kW (100 hp), 1800 rpm motors initially installed at pump mill sump.
Pump (replacement)	One of initial pumps replaced with a 6 by 6 by 17 PWOV, CISS pump to include pipe sections from pump to motor mounting plate; cutlass rubber shaft bearing for water lubrication; rated at 122.2 L/s (1940 gpm) at 38.4 m (126 ft) TDH; would pass 70 mm (2-¾ in.) solids.

Bleach Plant Sump (See Figure A-11)—

Function	To receive wastewater from bleach plant and chlorine dioxide generator system; pumps this wastewater to WWTP.
Sump dimensions	See Figure A-11
Construction	Reinforced concrete
Bar screen	None; no screening at the bleach plant sump

Pumps and Motors

Motors (initial)	Two vertical motors VHS, TEFC, non-overloading with 1.15 SF, 3/60/575 volts, 44.8 kW (60 hp), 1800 rpm.
Pumps (initial)	Two vertical turbine pumps, Model No. 14HH-276, all 317 stainless steel, rated at 154 L/s (2450 gpm) at 14.6 m (48 ft) TDH; would pass 25.4 mm (1 in.) solids.
Motor (replacement)	One of initial motors replaced with vertical in-line motor with high thrust, 1900 kg (4200 lb), Class F insulation, TKFC, P-base vertical solid shaft, to include "Chemical-Severe Duty" features; with capability (duplex bearings) frame, 404 hp, bearing life 2 years, B-10 rating; 44.8 kW (60 hp), 1200 rpm, 1.15 SF, 3/60/575 volts.
Pump (replacement)	One of initial pumps replaced with 10 by 8 by 17 SPWOV 317 SS pump with column, spider and hardware to be 316 SS epoxy coated; cutlass rubber shaft bearings for water lubrication; 316 SS epoxy coated mounting plate; rated at 9.3 m^3/m (2450 gpm) at 18.6 m (61 ft) TDH; would pass 70 mm (2-¾ in.) solids.

Kiln Area Sump (See Figure A-12)—

Function	To receive wastewater, overflows and spills from boiler feedwater plant, power plant general, maintenance and stores, lime mud storage, bark bin, wet end lime kiln, white liquor tank and new strong waste tank; pumps waste water, overflows and spills to WWTP.
Sump dimensions	See Figure A-12
Construction	Reinforced concrete
Bar screen	Identical to bar screen described for caustic sump except length is 5.4 m (17 ft 9¼ in.) sprocket center to sprocket center.

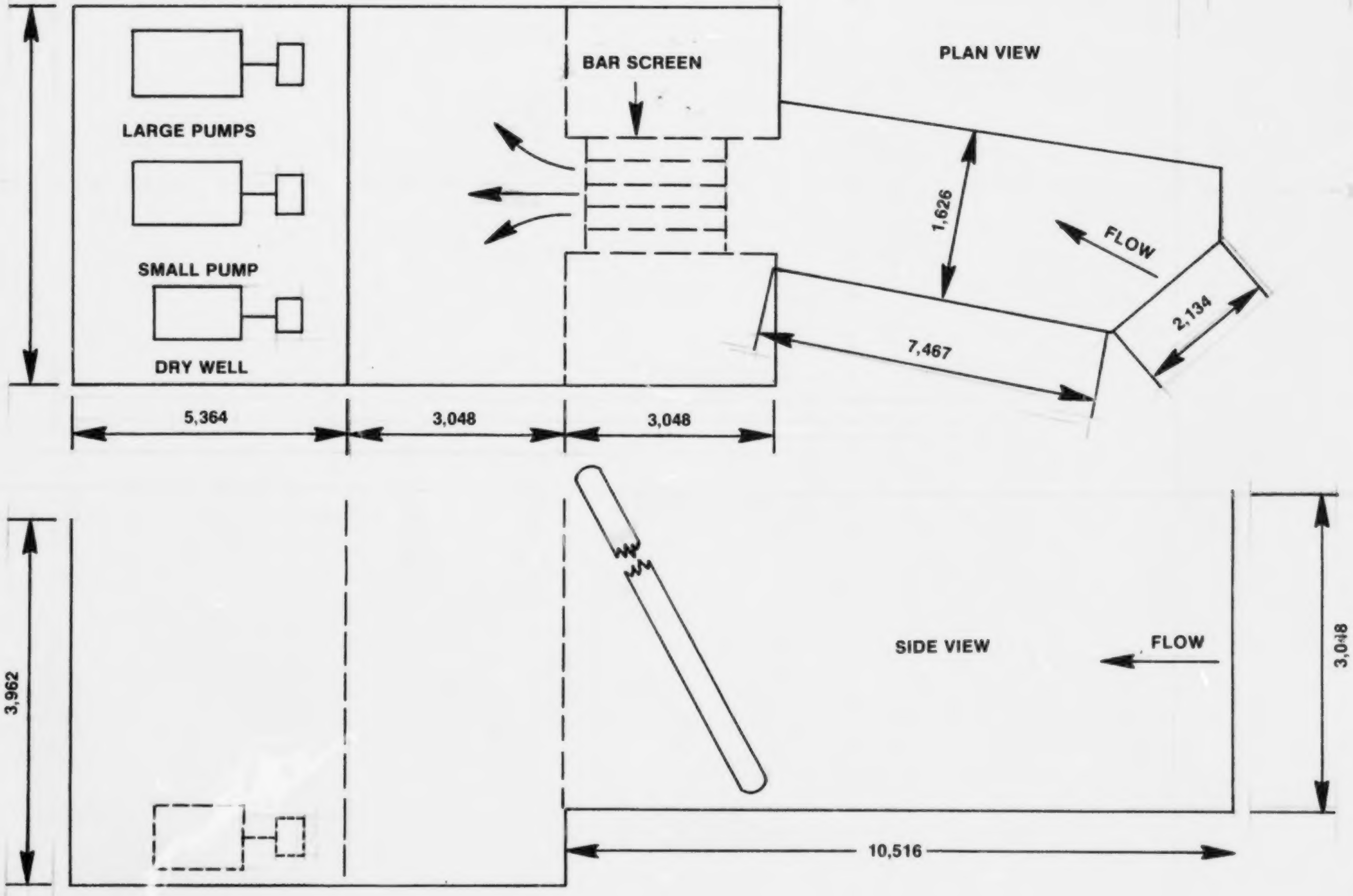


Figure A—12. Schematic view of kiln area sump. (Dimensions given in millimeters.)

Pumps and Motors

Motors (initial)	(For two larger pumps)—44.8 kW (60 hp), 1800 rpm, 3/60/575 volts, 405 U frame (motors flooded and burned out June 1978)
Motors (replacement)	(For above taken from inventory of spares)— (1) Model No. 2G7163, 44.8 kW (60 hp), 1780 rpm, 3/60) 575 volts, Code G, 1.0 SF, continuous duty 65°C (149°F), Class B insulation, serial No. 1E-73, 405 U frame; (2) Model No. 127, type RGZ, 44.8 kW (60 hp), 1770 rpm, 3/60/575 volts, code G, 1.0 SF, continuous duty 40°C (104°F), Class A insulation, NEMA Design B, serial No. 1-5106-08297-1-2, 405 U frame.
Pumps (two large)	Model No. 317 S, 4 by 6-14, 63 L/s (1000 gpm) at 40 m (131 ft) TDH, cast iron casing and side plate, 316 SS impeller and shaft sleeve, SAE 4140 shaft, Durametallic Type RO E55 O with 316 SS flush gland, grease lubrication, 150 cm by 350 mm (6 in. by 10 in.) cast iron suction piece, 316 SS suction wear plate, casing drain, suction and discharge gauge connection, cast iron baseplate sized for a 444 U frame motor, Falk 60T31 coupling and coupling guard.
Motor (initial)	(For one smaller pump)—29.8 kW (40 hp), 1800 rpm, 3/60/575 volts, 364 U frame motor (motor flooded and burned out June 1978)
Motor (replacement)	(For one smaller pump)—Model No. SF1.0, 29.8 kW (40 hp), 1800 rpm, 3/60/550 volts, Type E, design B, code F, continuous duty 55°C (131°F), serial no. P-222199.
Pump (one small)	Model No. 317 SS, 3 by 6-14, 44 L/s (700 gpm) at 42.1 m (138 ft) TDH, construction larger pumps described above except with 150 mm by 200 mm (6 in. by 8 in.) cast iron suction piece, cast iron base plate sized for a 365 U frame motor and Falk 50T31 coupling and coupling guard.

Strong Waste Storage Tanks and Control Systems—

Figure A-7 shows the basic flow diagram of the wastewater collection system. Strong waste storage tanks are shown at the bottom of the diagram just above the primary clarifier of the WWTP. Initially, only the caustic strong waste tank was designed into the system. Since then the kiln strong waste tank (a converted caustic storage tank) has been piped as a part of the system along with the kiln area sump described above under "kiln area sump". Both are described below as essential parts of the overall waste water collection system to smooth inevitable surges of strong waste to the WWTP.

Caustic Strong Waste Tank—

Function	To receive wastewater from caustic sump when pH at that sump is greater than 10 or less than 4; initially received spills or other recoverable streams from "boil-out" tank but such streams now return to chemical recovery.
Capacity	1516 m³ (400,000 gallons)
Dimensions	Diameter—15.3 m (50 ft)
	Height—9.2 m (30 ft)
Construction	A-283 Grade C carbon steel, 9.5 mm (⅜ in.) thick first ring, 2.34 m (7 ft 8 in.) high, remainder 8 mm (5/16 in.)

Control system	Continuous pH measurement at caustic sump opens control valve to caustic strong waste tank, closes control valve to wastewater line when pH at caustic sump exceeds 10; same action when pH drops below 4; reverse order operation when pH returns to range 4 to 10; manual override switch at lime kiln; tank level control indicator in WWTP control room.

Kiln Strong Waste Tank—

Function	To receive wastewater and collected spills from kiln area sump when pH at that sump is greater than 11 or less than 5. Meters strong waste to WWTP at reduced rate of 12.6 L/s (200 gpm)
Capacity	1622 m^3 (428,000 gallons) tank alone
Dimensions	Diameter—16.5 m (54 ft)
	Height—7.6 m (25 ft)
Construction	Carbon steel, concrete base and bottom; tank diked with creosoted bulkhead, concrete bottom, dirt backfill.
Control system	Flow to kiln area sump sampled and diverted to continuous pH monitor; full flow diverted to kiln strong waste tank when pH is greater than 11 or less than 5; tank overflow at top of straight wall to diked area, from there by gravity back to sump; 12.6 L/s (200 gpm) sumped continuously from kiln strong waste tank to WWTP.

Primary Clarifier System

Function	To receive composite waste water flow from all mill lift stations (sumps) and from both strong waste tanks; removes 70 to 75 percent settleable solids; rakes settled sludge to center and pumps it to sludge thickener, or directly, to be mixed with settled sludge from the secondary clarifier.
Construction	Reinforced concrete outside walls and bottom; bottom rake and plows of carbon steel skimmer of carbon steel.
Type	Center feed, peripheral overflow
Connected horsepower	3.7 kW (5 hp)
Dimensions	Diameter—38.1 m (125 ft)
	Side wall depth—3.1 m (10 ft)
	Bottom slope—25 mm (1 in.) in 305 mm (12 in.)
Load rate	62.7 $m^3/m^2/d$ (1328 $gal/ft^2/day$) at 61,800 m^3/d (16.3 mgd)
Type rake	Two long and two short arm rake equipped with blades to scrap primary sludge from outer area and inner area, respectively; equipped with blades to scrape outer area once and inner area twice per revolution at 4 RPH; also equipped with blades to scape 3.1 m (10 ft) annular trough around center well.
Torque limits	Alarms above set point of 48 percent; automatic shutdown at 88 percent.

Sludge discharge	Two primary sludge withdrawal points in annular trough, 180° apart.
Primary sludge pumps	Two centrifugal horizontal pumps with variable speed drives, rated at 14.2 L/s (225 gpm) at 9.2 m (39 ft) TDH
Backflush pump	Centrifugal horizontal pump rated at 20.5 L/s (325 gpm) at 45.8 (150 ft) TDH
Scum tank	1.14 m^3 (300 gallon) carbon steel tank; receives gravity drainage from dumpster box which collects plastics and other floating trash removed from primary clarifier by scum skimmer.
Scum pump	Centrifugal horizontal pump rated at 3.15 L/s (50 gpm) at 8.54 m (28 ft) TDH.

Cooling Tower System

Function	To cool total wastewater flow from primary clarifier before it enters the UNOX bioreactor.
Overall dimensions	Width—12.8 m (41 ft 11 in.) Length—15.1 m (49 ft 4 in.) Height—10.6 m (34 ft 10 in.)
Type	Three cell, mechanical induced draft, cross-flow tower complete with water distribution box and piping, water distribution basin, mist eliminators and three 44.8 kW (60 hp) fans. Each cell has a full width, full height transverse partition wall between fan cells so each cell may be independently operated.
Operating temperature range	62.8°C (145°F) to 35°C (95°F) with a design wet bulb of 25.6°C (78°F)
Cooling tower sump pumps	Three horizontal centrifugal pumps rated at 359 L/s (5700 gpm) at 16.8 m (55 ft) TDH. Two pumps can pump peak waste flow; third pump is a spare.

Nutrient and Defoamer Systems

Function	To provide storage for and additive rate control equipment for anhydrous ammonia, phosphoric acid (initially 75 percent; now 35 percent) and two types of defoamers; feeds the two nutrients to the wastewater stream between the cooling tower and bioreactor; feeds one defoamer into the wastewater stream ahead of the cooling tower and a second defoamer after the bioreactor.
Nutrient Storage— Anhydrous ammonia	Horizontal cylindrical tank, carbon steel, rubber-lined; capacity—56.9 m^3 (15,000 gallons), operates at 689 kPa (100 psig) with diurnal and seasonal variations.
Add-rate control system	(Initial) liquid ammonia vaporizer, steam heated, capacity 80.3 kg/hr (177 lb/hr) ammonia gas with water ejector 7.56 L/s (120 gpm) at 345 kPa (50 psig) to pull ammonia through ammoniator; (present) rotameter for liquid anhydrous ammonia, rate controlled by needle valve, liquid ammonia vaporized under ambient atmospheric conditions in 61 m (200 ft) pipeline, 76 mm (3 in.) diameter to cooled wastewater splitter box after cooling tower.

Phosphoric acid	Vertical, cylindrical FRP tank, capacity 19 m^3 (500 gallons)
Add-rate control system	Two positive displacement metering pumps, manually variable output of 0 to 45 L/hr (0 to 12 gph).
Defoamer storage	(Initial) one vertical cylindrical tank, capacity 24.6 m^3 (6500 gallons); (present) two vertical cylindrical tanks, each with capacity 24.6 m^3 (6500 gallons), insulated sides and top, one steam heated, one electrically heated to maintain temperature of 21.1°C (70°F); one tank stores defoamer for high pH situations, both tanks have agitators.
Pumps	Two per tank, positive displacement metering pumps, 0 to 45.5 L/hr (0 to 12 gph).
UNOX Bioreactor	
Function	To provide retention time in totally enclosed vessel for bio-degradation of oxygen-consuming organic materials in combined mill wastewater; provides mass transfer of high-purity gaseous oxygen to liquid phase wastewater by use of surface aerators; provides safety features to detect hydrocarbons and to purge gas space rapidly with air when necessary; maintains substantial DO concentration in wastewaters to provide aerobic bidegradation of wastes.
Dimensions	(See Figure A-13)
Construction	Reinforced concrete basin with gas-tight concrete top.
Number of trains	Two in parallel
Stages per train	Three
Connected power	Stage No. 1, each train-150 kW (200 hp)
	Stage No. 2, each train—55 kW (75 hp)
	Stage No. 3, each train—45 kW (60 hp)
	Total (2 trains)--500 kW (670 hp)
Motors	Stage No. 1, each train—two 445 U motors, 3/60/550 volt, 1180 rpm TEFC
	Stage No. 2, each train—one 44 U motor, 3/60/550 volt, 1750 rpm, TEFC
	Stage No. 3, each train—one 405 U motor, 3/60/550 volt, 1750 rpm, TEFC
Aerators—	
Function	To agitate liquid phase in bioreactor and creates gas-liquid interfacial area for oxygen transfer to liquid phase (See Table A-1 for pertinent description).

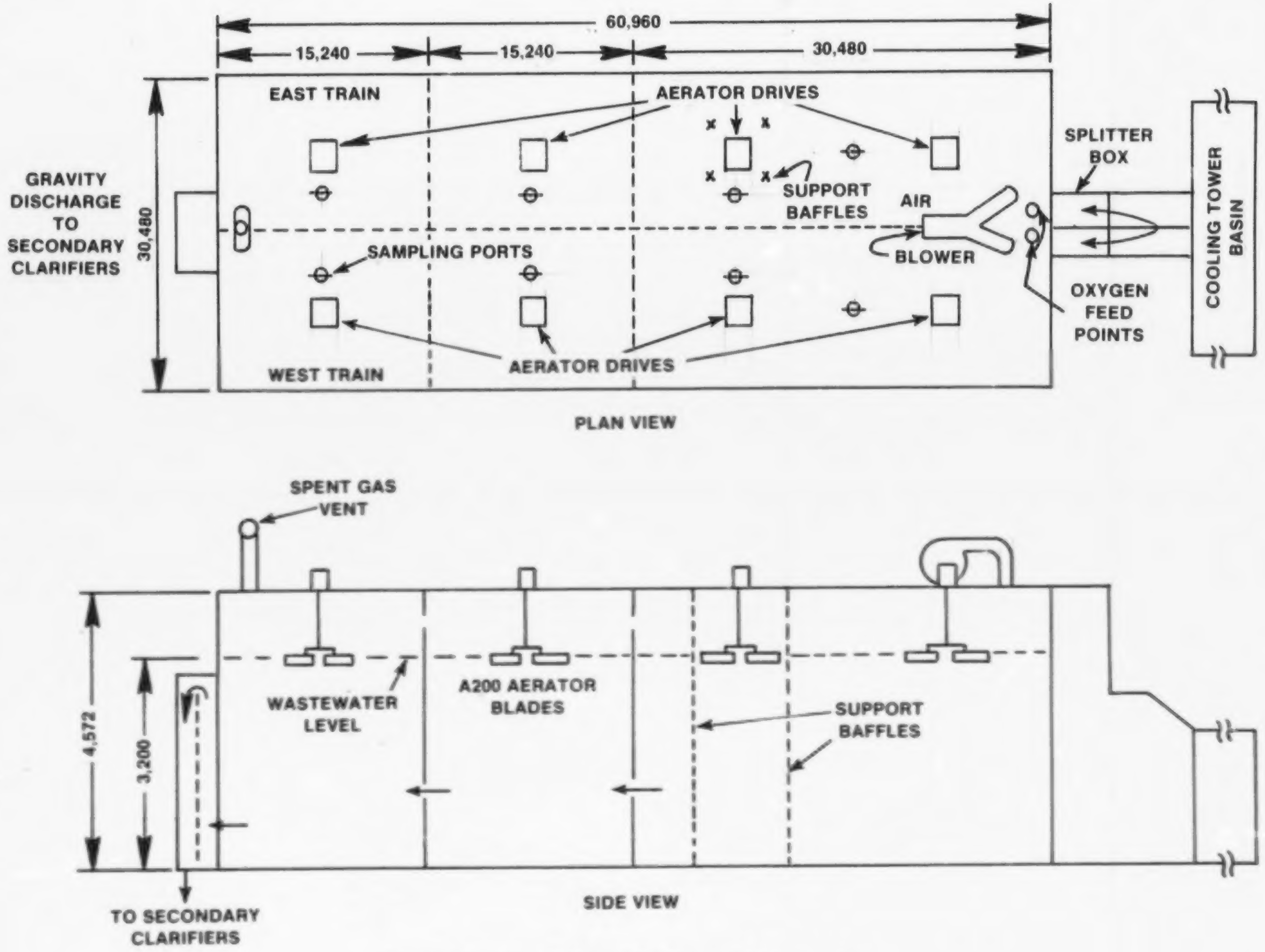


Figure A—13. Schematic views of bioreactor.
(Dimensions given in millimeters.)

TABLE A-1. AERATOR CHARACTERISTICS BY STAGE

Characteristic	Stage no. 1	Stage no. 2	Stage no. 3
Shaft diameter, mm (in.)	140 (5½)	140 (5½)	127 (5)
Blade width, mm (in.)	387 (15-¼)	368 (14-½)	327 12-⅞)
Blade thickness, mm (in.)	25.4 (1)	22.2 (⅞)	22.2 (⅞)
Number of blades/aerator	4	4	4
Type of blade	A200	A200	A200
Number of aerators/stage	2	1	1
Overall diameter, m (in.)	2.67 (105)	2.54 (100)	2.26 (89)
Vertical downward load, kg (lb)	9,843 (21,700)	7,439 (16,400)	5,579 (12,300)
Torque, cm-kg (in.-lb)	357,160 (310,000)	215,050 (187,000)	181,700 (158,000)
Bending moment, cm-kg (in.-lb)	698,050 (607,000)	552,000 (480,000)	489,900 (426,000)
Impeller, rpm	45	45	45

Safety System—

Function	To detect presence of hydrocarbons in gas space above wastewater in bioreactor; at certain fractions of lower explosive limit (LEL) starts purge air blower and stops oxygen flow and then shuts down aerators.
Purge air	oxygen flow stopped—25 percent of LEL
Purge time	Entire bioreactor—2 min
Capacity	Purge air blower—5660 m³/hr (200,000 cfh)
Aerators shutdown	50 percent of LEL

Secondary Clarifier System

Function	Receives biodegraded wastewater from bioreactor; settles, thickens and recycles biomass back to bioreactor to maintain proper MLSS levels; diverts excess biomass, or sludge, to thickener or directly to vat of dewatering filters; maintains sludge blanket depth in clarifier; discharges overflowing clarified effluent to pipe leading to river diffuser.
Number of clarifiers	2
Construction	Reinforced concrete, outer race later elevated by 305 mm (1 ft 2 in.) of brick.
Type of clarifier	Peripheral feed, peripheral overflow
Diameter	41.2 m (135 f)
Side wall depth	4.3 m (14 ft)
Bottom slope	5.2 mm/meter (1/16 in./1 ft)
Load rate	24.5 m^3/m^2 day (600 gpd/ft^2) at 6177 m^3/day (16.3 mgd)
Type of rake	Single tapered hollow steel header with series of variable size inlet orifices over length of rake.
Rake rotation	2.2 rph
Motor drive on rake	0.56 kW (¾ hp)
Secondary sludge recycle pumps	Three pumps for two clarifiers, centrifugal horizontal pumps each rated at 132.3 L/s (2100 gpm)
Normal sludge blanket depth	1.53 m (5 ft)
Design sludge underflow solids	2.5 to 3 percent

Sludge Thickner

Function	Receives sludge underflow from primary and secondary clarifiers for thickening before sludge dewatering.
Construction	Painted steel
Capacity	273 m^3 (72,000 gal)
Diameter	10.7 m (35 ft)
Sidewall depth	3.1 m (10 ft)
Slope	16.7 mm/meter (2 in. by 1 ft)
Sludge pumps and motors	Two centrifugal combined sludge pumps rated at 23.6 L/s (375 gpm), each fitted with variable speed drive and 14.9 kW (20 hp) motor.

Sludge Removal System—

Function	To thicken mixture of primary and secondary sludge; moves thickened mixture to bottom center cone for removal to vat of dewatering filters.
Type of rake	Two arms with vertical pickets on top of arms for thickening; steel plates with brass squeegees on bottom of rake.
Drive	(initial) 0.56 kW (¾ hp), 3/60/575 volts, TEFC, 1800 rpm reduced to rake rpm of 0.042; (present 0.75 kW (1 hp), 3/60/575 volts, TEFC, 1725 rpm input to variable speed system with maximum output to chain drive of 68 rpm and minimum of 13.6 rpm giving variable rake rotation of 2.5 rph to 0.5 rph.
Sludge influent	0.82 m (2.7 ft) down from top into side of 1.22 m (4 ft) high by 2.44 m (8 ft) diameter center well.
Sludge effluent	Withdrawn at center cone at bottom.
Top overflow	Returned to primary clarifier inlet
Detention time	16 hr

Sludge Dewatering Vacuum Filters

Function	To receive combined sludge from the thickener, or directly from the primary and seconedary clarifier underflow when the concentration of these sludge streams is at 3.5 to 4.0 percent and 1.5 to 2.5 percent, respectively.
Size and loading of filter drums	Filter drums—diameter—3.1 m (10 ft)
	Width—2.5 m (8 ft)
Design loading	43.9 kg/m²/hr (9 lb/hr/ft²) at 3 percent influent sludge concentration
Filter cover	Coil springs of 304 stainless steel

Auxiliary Equipment for Filters—

Agitator	One in each filter vat, each driven by 2.2 kW (3 hp) motor.
Drum drive	0.37 kW (½ hp) motor on each drum
Filtrate receiver	One for each drum, diameter 1.07 m (42 in.), height 2.9 m (9 ft 6 in.)
Vacuum pump	Capacity of 48.1 m³/min (1700 cfm) at 330 mm (13 in.) mercury vacuum driven by 56 kW (75 hp) motors
Filtrate pump	Centrifugal rated at 18.9 L/s (300 gpm), driven by 3.73 kW (5 hp) motor.
Sludge chute	Centrally fixed diverter gate to load sludge alternately into two trucks standing side by side.

Treated Effluent Discharge Pipe

Function	Initially measured treated effluent flow by Parshall flume; conveyed stream to concrete foam pit. thence to flow control valve in grate-covered concrete enclosure at top of river bank, down river bank through concrete reinforced pipe; present system is the same except flow at WWTP is now measured after primary clarifier and flow control valve mode of control has changed.
Pipe diameter	914 mm (36 in.)
Pipe length	305 m (1000 ft) from Parshall flume to river
Type and size of valve	Butterfly, 457 mm (18 in.) body, elastomer lined; disc of steel, shaft of 304 SS, full open capacity 788 L/s (12,500 gpm)
Valve opening control	Operates to control flow from level in upstream foam control vat, said level sensed by bubble tube; control system set so that valve will fail in open position.

Treated Effluent Submerged Diffuser

Function	To receive treated effluent from WWTP through underground pipe with control valve at top of river bank; diffuses treated effluent into receiving river water at depth of 7.63 to 9.15 m (25 to 30 ft).
Diffuser diameter	813 mm (32 in.) O.D.
Diffuser length	91.5 m (300 ft)
Construction	25 mm (1 in.) wall thickness, polyethylene pipe
Nozzle distribution	76 mm (3 in.) diameter PE nozzles 152 mm (6 in.) long, fitted with 102 mm (4 in.) PE 90° els facing up and down river alternately on adjacent nozzles; nozzles located on top of diffuser pipe spaced 1.83 m (6 ft) apart.
Nozzles installed	Over approximately 91.5 m (300 ft) length of large pipe; the nozzle section located in east half of river section beginning near river center.
Diffuser pipe support	254 mm (10 in.) O.D. pipe pilings, 15.3 m (50 ft) long; special yoke of 152 mm (6 in.) steel pipe and shaped wood secured to top of pipe pilings to anchor diffuser pipe; total of 12 pipe pilings over 183 m (600 ft) length of diffuser pipe.

Sludge Disposal System

Function	To receive dewatered sludge from chute of vacuum filters; hauls sludge to landfill area 14 miles one way; mixes bark and sludge when necessary for landfilling; spreads sludge or sludge and bark mixture on landfill; covers sludge or mixture with earth periodically.
Trucks	Two welded body 10-wheel, capacity 11.45 m^3 (15 yd^3), normal load 9.54 m^3 (12.5 yd^3); variety of mill trucks haul bark separately when needed at landfill.
Loads hauled	Twelve per day during average 18 hour period, 7 days per week.
Equipment at landfill site	Two dozers, each with front blade and 914 mm (36 in.) wide tracks; one dozer is lighter weight than the other for travel over virgin sludge mixed with earth or bark; the heavier dozer moves mixed sludge and digs out earth for coverage.

BLACK LIQUOR OXIDATION SYSTEM (BLO)

Function	To oxidize sodium sulfide in weak black liquor to sodium thiosulfate using high purity oxygen (99.5 percent) in simple pipeline reactor.

Oxygen Pipeline From Compressor

Function	To convey high-purity oxygen from compressor, through orifice section to flow control valve and thence to the sparge section of the in-line BLO reactor.
Dimensions	Diameter—102 mm (4 in. I.D.)
	Length—153 m (500 ft)
Construction	(Initial) low carbon steel (replacement before startup and present) 316 stainless steel

Flow Measurement and Control System—

Function	Measures and controls oxygen sparged into weak black liquor.
Type of flowmeter	Flange orifice
Orifice diameter	73 mm (2.875 in.)
Construction	316 stainless steel
Type of valve	Type 9500 butterfly
Valve size	76 mm (3 in.)
Construction	316 stainless steel
Control system operation	See following paragraph and Figure A-14

Weak Black Liquor Storage and Flow System (See Figure A-14)

Function	To store weak black liquor ahead of BLO sparge and reactor sections of pipeline; controls flow of weak black liquor to evaporators by set point pressure at header feeding evaporators.
Liquor storage tank dimensions	Height—7.6 m (25 ft)
	Diameter—14 m (46 ft)
Construction	Steel
Pumps	Two centrifugal in parallel (or one as spare) PWO-8 by 6-17 with 362 mm (14¼ in.) stainless steel impellers; cast iron casing; each rated at 139 L/s (2200 gpm) at 51.9 m (170 ft) head or 183 L/s (2900 gpm) at 44.2 m (145 ft) head.
Motors	Two squirrel cage induction motors, 112 kW (150 hp), 1800 rpm frame M505, TEFC, 3/160/575

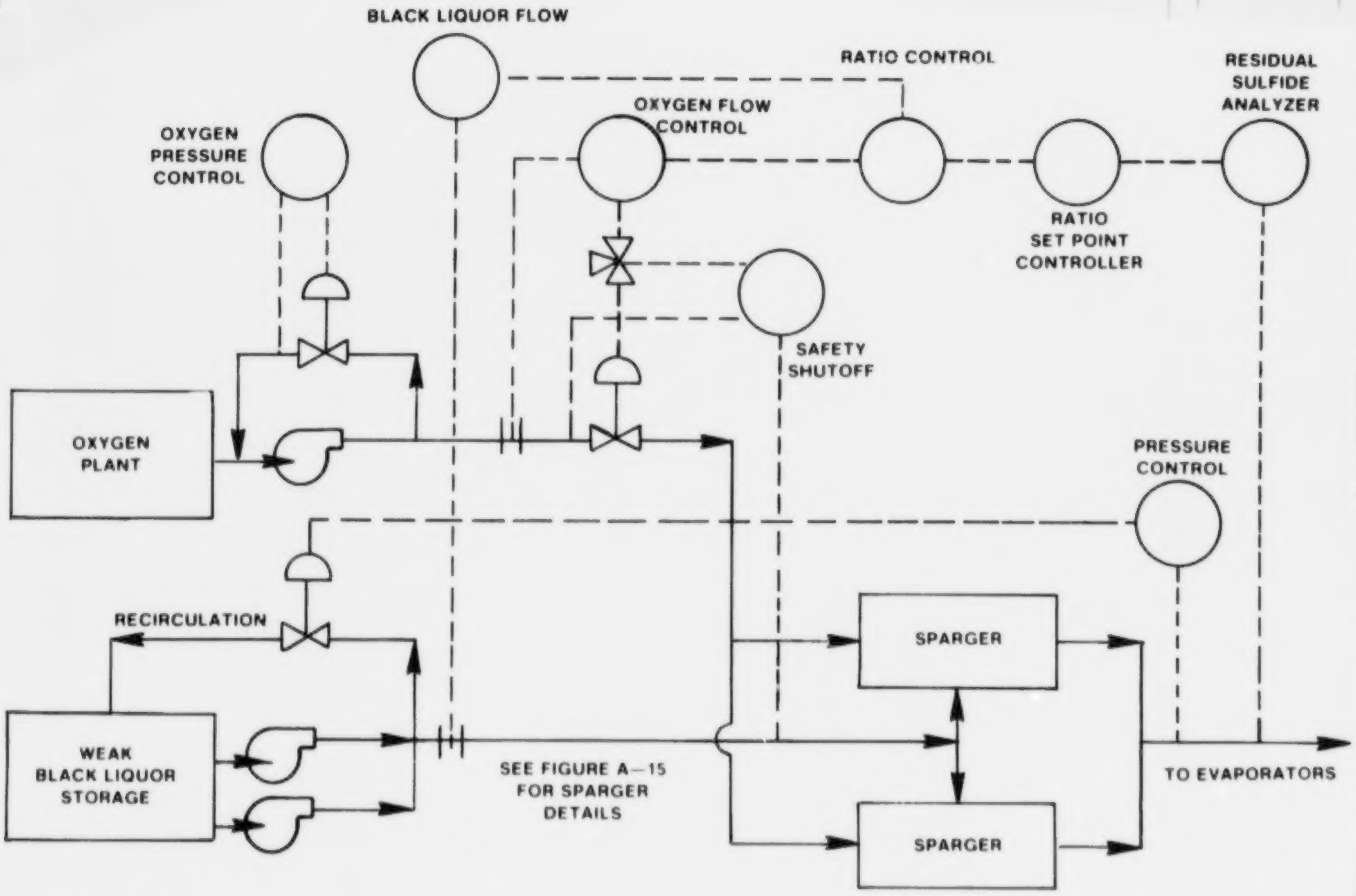


Figure A—14. Black liquor oxidation system.

Bypass Valve	102 mm (4 in.) butterfly in 152 mm (6 in.) recirculation line back to the weak black liquor storage tank.

Oxygen Sparger Section

Function	To receive high-purity oxygen under pressure in drilled sparge pipe (initially two pipes) to provide intimate contact with weak black liquor flowing in larger "tee" section surrounding oxygen sparger pipe (see Figure A-15, this Appendix)
Sparge Pipes	Length—0.8 m (2 ft 7 in.) inside liquor "tee"
	Diameter—38 mm (1½ in.) schedule 40
	Construction—304 L stainless steel
	Sparge hole diameter—1.6 mm (1/16 in.)
	No. sparge holes/row—93
	No. of rows—4
	Spacing of holes—3.2 mm (⅛ in.) center to center
	Length of hole row—292 mm (11.5 in.) starting; 406 mm (16 in.) and ending; 114mm (4.5 in.) from capped end of sparge pipe
	No. of sparge pipes—(initially) two; (presently) one
Sparge pipe connected to	102 mm (4 in.) oxygen supply line through 102 mm (4 in.) manually operated shut off valve, 102 mm (4 in.) check valve to prevent liquor back up into oxygen line and 38 mm (1½ in.) manually operated shutoff valve.
Weak black liquor "tee"	Modified 203 mm (8 in.) standard tee (see Figure A-15) with 254 mm by 203 mm (10 in. by 8 in.) reducer to connect to existing 254 mm (10 in.) weak black liquor 254 mm (10 in.) manually operated shut off valve; discharge through existing 203 mm (8 in.) manually operated shut off valve; by-pass line provided (not shown in Figure A-15) around sparge section for weak black liquor.

Shear Rate Reactor Section

Function	To provide appropriate shear rate in weak black liquor pipeline to achieve essentially complete reaction between injected oxygen and sodium sulfide.
Reactor dimensions	Length—30.5 m (100 ft)
	Diameter—203 mm (8 in.)
Construction	Carbon steel

Residual Sulfide Measurement System

Function	To receive continuous sample of oxidized weak black liquor; senses and measures residual sulfide concentration; transmits signal to oxygen flow controller to adjust flow up or down as needed at constant liquor flow.

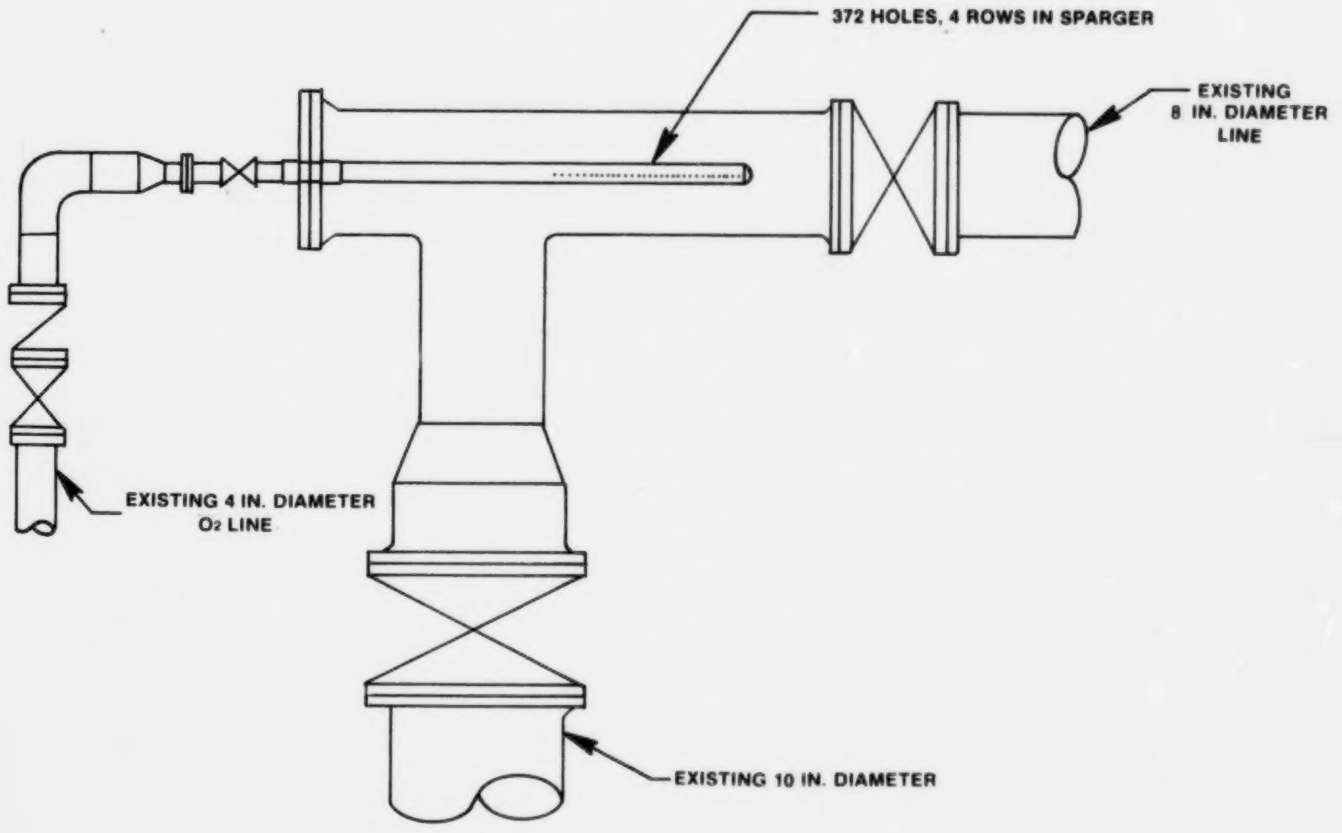

372 HOLES, 4 ROWS IN SPARGER
EXISTING 8 IN. DIAMETER LINE
EXISTING 4 IN. DIAMETER O_2 LINE
EXISTING 10 IN. DIAMETER

Figure A—15. Schematic view of oxygen sparger section.

Type	"Electrofact" probe with potassium chloride (KCl) solution as reference in conjunction with a silver billet electrode; (initial) KCl vial exposed directly to hot black liquor; (present) KCl vial covered by heat-shrink tubing to prolong life.

Overall Oxygen—Black Liquor—Sulfide Control System

Function	To control flow of oxygen to BLO system as weak black flow or residual sulfide content in oxidized black liquor changes (Figure A-14 shows relationship of the various parts of the system; Table A-2 lists the points controlled and the generic type of signal and controller; operation of the system is described in Appendix C).

TABLE A-2. GENERIC DESCRIPTION OF THE OVERALL BLO CONTROL SYSTEM

Point controlled	Type of signal	Type of controller
Oxygen pressure	Pressure	Pneumatic
Oxygen flow	Orifice/DP Trans./I/P	Electronic
Black liquor flow	Orifice/DP Trans./I/P	Input to electronic
Ratio controller	Electronic from black liquor flow and residual sulfide	Electronic
Analyzer controller	ORP/mA transmitter	Electronic
Black liquor pressure	Pressure	Pneumatic
Safety shutoff	Oxygen pressure and black liquor DP transmitter	Pneumatic switch

APPENDIX B

EXPERIMENTAL PROCEDURES

The procedures used in developing report data are summarized below for the following systems:

(1) In-Process Changes

(2) Bleaching System

(3) Air Separation Plant

(4) Wastewater Treatment Plant

(5) Black Liquor Oxidation System

IN-PROCESS CHANGES

Total Reduced Sulfur (TRS) Compounds in Condensate Streams

In this analysis, a known quantity of iodine is added to an aliquot of sample. The iodine consumes the TRS by the reaction

$$I_2 + S^{=} \longrightarrow 2I^- + S$$

The remaining excess I_2 is titrated with sodium thiosulfate under acid conditions via the reaction

$$I_2 + 2S_2O_3 \longrightarrow 2I^- + S_4O_6$$

The TRS is then calculated as the difference between the I_2 added and the excess I_2 as follows:

$$\text{normality TRS (mL sample)} = \text{normality } I_2 \text{ (mL } I_2\text{)} - \text{normality thio (mL thio)}$$

$$\therefore \frac{\text{MG TRS AS } H_2S}{\text{liter}} = \left[\frac{\text{normality } I_2 \text{ (mL } I_2\text{)} - \text{normality thio (mL thio)}}{\text{mL sample}}\right] \times 1700$$

If a 25-mL sample is used, then the stripper inlet samples will require 10 mL of 0.1N iodine and the stripper outlet samples will require 10 mL of 0.01N iodine. The 0.01N sodium thiosulfate is used as the titrant to a starch end point. The procedure is as follows:

(1) Pipette 25 mL-sample into a 125-mL erlenmeyer flask

(2) Add iodine and swirl

(3) Add 10 mL of 8N H_2SO_4

(4) Titrate to starch endpoint with 0.01N sodium thiosulfate until the color does not return for thirty seconds.

In the event that the amounts of iodine given are insufficient to produce a permanent iodine color in the sample after step (2), then add another similar aliquot of iodine and proceed to step (3).

Methanol

Methanol concentration is determined according to following procedure:

(1) Filter sample with filter paper to remove turbidity

(2) Put 5-mL sample in 50-mL flask

(3) Add 15 drops H_2PO_4

(4) Add 25 drops $KMnO_4$

(5) Swirl occasionally for 10 min

(6) Add $NaHSO_3$ dropwise till colorless

(7) Place in ice bath

(8) Add 25 mL cold concentrated H_2SO_4

(9) Add 20 drops chromotropic acid reagent

(10) Swirl at 60°C for 15 min

(11) Cool in ice bath, add water almost to mark

(12) Cool to room temperature, add water to mark

(13) Mix and run color at 605 mu, open filter (use blank to obtain zero absorbence).

(14) Determine CH_3OH concentration from calibration curve made from known mixtures of methanol and water.

BLEACHING SYSTEM

Procedures used to develop data for selection and evaluation of the oxygen bleach plant are divided into those used by CIL and those used by CCV. All reported data from references 15 and 41 (CIL reports) were by CIL procedures.

CIL Bleach Plant Procedures

(1) Final Brightness—determined on 4 g handsheets made from pulp soured with SO_2, and read on an Elrepho brightness meter using a Tappi filter (457 mu)

(2) Interstage Brightness—same as (1) above except without souring with SO_2.

(3) Brightness Reversion—loss in brightness measured by heating tabs from handsheets for 1 hour at 105° C in a forced air circulating oven.

(4) Viscosity—conducted according to Tappi Standard T230 using the efflux viscometer method.

(5) Kappa Number—conducted according to Tappi Standard T236M60.

(6) BOD_5—conducted according to CPPA H2P.

(7) Color—conducted according to Tappi Standard 620. Sample pH adjusted to 7 before color measurement made.

(8) Chlorides—according to procedure for chlorides under section entitled "CCV Bleach Plant Procedures".

(9) Pilot D/COD Sequence Bleaching Procedures

Bleaching Conditions—
Each experiment was conducted with sufficient pulp to yield a minimum of 15 to 20 g o.d. weight of the pulp for each set of conditions in the last stage. Certain special techniques are discussed below:

D/C Sequential Addition—
Chlorine dioxide was added to the preheated pulp slurry with stirring and allowed to react for two minutes. The required quantity of chlorine water was then added to yield a final consistency of 3 percent and a temperature of 50° C. The total reaction time was fixed at 10 min but the total available chlorine and the percentage of equivalent chlorine dioxide were varied.

Oxygen Stage—
Where applicable, magnesium complex was incorporated into the pulp by slurrying the pulp at 3 percent consistency with the calculated amounts of magnesium and complex. The pulp was filtered and pressed to approximately 40 percent consistency before crumbing. The requisite quantity of sodium hydroxide and dilution water were slowly added to the pulp in a Hobart mixer. To ensure thorough distribution of the chemical, the pulp was manually pounded, after which the pulp was crumbed and reacted. Oxygenation was carried out for 30 min at a pressure of 620 kPa (90 psig) and a temperature of 120° C.

Effluent Carry-Over—

Following the oxygen stage, the pulp was diluted to 4.0 percent consistency and after the pH was measured, the pulp was filtered and the filtrate collected. The pulp was then washed well before samples were taken for pulp evaluation. Effluent from the second stage was carried over into the third stage in order to simulate washer efficiency in a typical mill bleachery and to control the pH during the chlorine dioxide stage. The amount of carryover was estimated by adding the effluent to a mixture of the required dilution water and chlorine dioxide until the consistency was 6 percent and pH was in the range 6.5 to 7.0. This amount of carryover, together with the pulp and dilution water was heated in order that the temperature after addition of chlorine dioxide solution was at the desired reaction temperature.

Chlorine Dioxide Stage—

Chlorine dioxide was mixed with pulp at 6 percent, and reacted for 3 to 7 hours at temperatures from 75 to 85°C.

CCV Bleach Plant Procedures

Determinations of the permanganate number were conducted according to the following procedure:

(1) Make K-No. hand sheet by blending in blender about 5 g of pulp.

(2) Make hand sheet on sheet mold.

(3) Dry sheet on hot plate (do not burn).

(4) Place hand sheet in oven (220°F and dry for at least 30 minutes or more).

(5) Weigh on balance 0.500 g (± 0.001 g) of the dry hand sheet.

(6) Place sample back in oven for 1 to 2 min and reweigh and record exact weight.

(7) Place pulp sample in blender about half full of tap water.

(8) Blend until pulp is completely dispersed.

(9) Pour contents of blender into a battery jar, rinse blender jar and lid and pour into battery jar.

(10) Fill battery jar to line (1140 mL) with tap water.

(11) Place on stirring apparatus and lower stirrer to within 127 mm of bottom jar and start stirrer. (Move the jar so that the stirrer is to one side of the jar to avoid forming a vortex great enough to draw air into the solution.)

(12) Add 20 mL of 8N sulfuric acid (H_2SO_4) from the automatic pipet.

(13) Add 40 mL of 0.1N potassium permanganate ($KMnO_4$) from the automatic pipet and start the stop watch after 20 mL of solution has gone into the jar as indicated by a mark in the middle of the pipet.

(14) Time for exactly 5 minutes.

(15) At the end of 5 minutes add immediately 25 mL of potassium iodide (KI) solution from the automatic dispenser squeeze bottle.

(16) Immediately start titrating with 0.1N sodium thiosulfate ($Na_2S_2O_3$) from the burette to a light straw color.

(17) Add starch solution (the solution will turn blue) and continue to titrate until the solution turns colorless.

(18) Record mL of thiosulfate used.

(19) Record temperature of the solution in the jar.

(20) Neglect return of the blue color.

(21) Calculation example:

Thiosulfate used - 22.2 mL

Temperature - 23° C

Permanganate added - 40.0 mL

Weight of sample used - 0.505 g

mL $KMnO_4$ consumed	Weight corrected to 0.500 g
$\begin{array}{r} 40.0 \\ \underline{-22.2} \\ 17.8 \end{array}$	$\frac{17.8 \times .5}{.505} = 17.6$

Correct sample to 1 g sample $\quad 17.6 \times 2 = 35.2$

If temperature is above 22° C, subtract chart value.

If temperature is below 22° C, add chart value.

Temperature corrected to 22° C from chart $\quad 35.2 - .5 = 34.7$

K-No. is 34.7

Determinations of viscosity were conducted to the following procedure:

Sample Preparation—

(1) Tear the equivalent of total 30 g from the bleached pulp sample (sample could have one or more sheets).

(2) Put these pieces in the Standard (British) Disintegrator and add 2 liters of water.

(3) Disintegrate for 60 counts.

(4) Take a 100 mL of the pulp slurry using 250 mL plastic beaker. Add about 100 mL acetone into the beaker and pour into a 90 mm Buchner funnel (use No. 4 filter paper).

(5) Apply the vacuum and add immediately acetone in excess. Repeat adding acetone at least one more time.

(6) After washing with acetone pull the vacuum for short period of time—about 20 to 30 seconds.

(7) If the sample was treated properly the pad should be fluffy and not compressed against the filter paper.

(8) Take pad from funnel and remove filter paper.

(9) Air dry the pad 24 hours or at least overnight (16 hours) in the humidity room.

(10) Weigh an amount of air dried pulp to equal 250 mg of oven dry pulp. This has been determined to be 269 mg of the pad prepared and dried as described above.

Viscosity Determination—

(1) Place 269 mg of the pulp into a clean dry 125 mL erlenmeyer flask, containing already 31.8 mm magnetic stirring bar.

(2) Add 50 mL 0.5 M CuEn solution from automatic pipet.

(3) Flush out the vapor space over the CuEn solution in the flask with dry nitrogen gas and put the stopper on the flask.

(4) Dissolve the pulp in CuEn by stirring on magnetic stirrer for 30 minutes.

(5) Load clean and dry viscometer in an inverted vertical position with the capillary side submerged in the solution.

(6) Apply suction to the other arm of the viscometer and fill both small bulbs on the capillary arm with the solution.

(7) Bring the liquid into the working capillary to the etched mark.

(8) After filling the pipet, rotate it to its normal vertical position and place in a constant temperature water bath controlled at 25.0 ± 0.1° C.

(9) The solution will reach the bath temperature in about 5 min.

(10) When this temperature is reached, determine the efflux time by drawing the solution above mark between the two bulbs and measuring the time required for the meniscus to pass from the mark between the two bulbs to the mark below the lower bulb.

(11) Make at least two efflux time determinations on each pipet; the results should check within ± 0.2 seconds. Use average time of two measurements within ± 0.2 seconds.

(12) Duplicates should be run on each sample.

(13) Calculate the viscosity V of the 0.5 percent pulp solution from the formula:

$$V = C\,t\,d$$

Where:
V = viscosity in centiposes
C = the pipet constant found
t = efflux time in seconds
d = the density of 0.5 percent of the pulp solution which is 1.052

GE brightness—According to instruction manual for Model S-4 Standard Brightness Tester and Colorimeter as manufactured by Diano Corporation and calibrated monthly with samples furnished by Institute of Paper Chemistry.

Valley beater test—Conducted according to Tappi Standard T-200 OS-70 with the following exception:

(1) Tap water with a conductivity of about 1000 micromhos/cm is used for beater dilution as well as freeness testing.

(2) Sample clearing was not performed because it was not necessary.

Sheetmaking procedure—According to Tappi Standard 205 OS-71 with the exception that the second pressing we did not separate the sheet and disk from the original couch blotter. This deviation helps prevent wrinkling when the sheets are dried in the rings.

Breaking length—According to Tappi Standard T-404.

Tear—According to Tappi Standard T-414 with the exception that the tear tester has a shallow cut out on the pendulum.

Burst—According to Tappi Standard T-403.

Bulk—According to Tappi Standard T-220-OS-71.

Opacity—According to Tappi Standard T-425.

Freeness—According to Tappi Standard T-227. All freeness measurements during beater testing done with tap water with a conductivity of about 1000 micromhos/cm.

Fiber length—According to Tappi Standard T-233 using Bauer-McNett type classifier.

Per[illegible]nt hardwood and percent pine—According to Tappi Standard T-401.

Dirt—According to Tappi Standard T-213 (shives are counted and reported separately).

Sodium (Na+)—Run with Beckman Model DU-2 Flame Spectrophotometer.

Shives—According to procedure outlined below: Shive counts are made by reflected light at about a 35° angle. Both sides of a 762 mm by 762 mm sheet are counted for number of shives. The total number of shives on both sides of the sheet are reported. Shives are identified by color and shape. Questionable shives are examined under a 10 times to 20 times power enlarger.

Chlorides—According to procedure outlined below:

Apparatus:

1 — Burrette, Automatic 0-10 or 0-25 mL

1 — Casserole, porcelain, 210 mL capacity

1 — Cylinder, graduated, 50 mL

1 — Dropper bottle, indicator type

1 — Stirring rod, glass

Chemicals Required:

Sulfuric acid, 1/50N

Silver nitrate, 0.0282N (1 mL = 1 mg Cl^-)

Phenolphthalein indicator

Potassium chromate indicator

Sodium hydroxide, 1N

Procedure:

(1) Filter sample if cloudy and measure 100 mL of water into porcelain dish. If chloride content is high ½ or ¼ sample size can be used.

(2) Add 2 drops of phenophthalein indicator. If no pink color appears add enough 1N NaOH to just turn solution pink.

(3) Add 1/50 H_2SO_4 till the pink color just disappears.

(4) Add 3 to 5 drops potassium chromate indicator to the sample.

(5) Fill burette with standard silver nitrate solution to the zero mark, making sure that the tip of the burette is filled with solution.

(6) Add silver nitrate through burette to the sample of water until yellow color of potassium chromate first turns to a slight pinkish tinge. This is the end point, not a brick red color that will develop if additional $AgNO_3$ is added.

(7) From the mL of $AgNO_3$ used subtract 0.2 mL and multiply by 10 and record as total ppm Cl^-. (0.2 mL is subtracted because this is the amount of $AgNO_3$ required to produce the endpoint in 100 mL of distilled H_2O.) Multiply by 20 if ½ sample is used, multiplying by 40 if ¼ of sample is used.

Suspended solids by sheet mold method—According to procedure outlined below:

(1) Collect a sample large enough to enable thorough mixing of fibers (4 L).

(2) Mix 4-liter sample thoroughly and pour into 2 liter graduated cylinder immediately.

(3) Pour 2 liter sample into sheet mold, open drain valve to form sheet. Open sheet mold.

(4) Place 3 to 4 dry blotters on wet sheet and roll with hand roller 4 or 5 times to help blotters absorb excess water.

(5) Remove carefully and place on hot plate with one blotter on top of sheet and dry.

(6) Observe sheet during drying time to prevent burning (changes weight) of handsheet.

(7) Weigh on balance to 2 decimal places.

(8) Dry weight of handsheet in grams times 500 equals mg/L fiber loss.

(9) Multiply value obtained in (8) by 1.02 and add 240 mg/L to calculate suspended solids level as measured by Standard Methods.

Note:

If use of volume other than two liters is necessary, concentration of suspended solids can be calculated from the following:

$$\frac{\text{Weight of dried solids in grams}}{\text{Volume of sample in liters}} \times 1000 = \text{mg/L suspended solids}$$

Pilot D/COD sequence bleaching procedures—

General Procedure

(1) Establish bleach trial parameters in terms of o.d. pulp, chemical levels, reaction temperatures, reaction duration, and reaction consistency.

(2) Determine percent solids of pulp to be bleached after running pulp through the fluffer.

  (a) Weigh wet sample, dry and weigh. (Forming handsheet and drying on hot plate is quickest method.)

  (b) Percent solids equals dry weight divided by wet weight.

(3) Establish chemical strength.

Cl_2 (chlorine water)

(a) Put 10 mL of 10 percent KI (potassium iodide) plus 10 mL of 10 percent acetic acid plus 100 mL distilled water into erlenmeyer flask.

(b) Pipette 10 mL of chlorine water sample into flask, being careful to hold pipette tip below liquid surface.

(c) Titrate with 0.1N $Na_2S_2O_3$ (sodium thiosulfate) until yellow, add starch to blue and continue titrating until colorless.

(d) Grams per liter Cl_2 = mL $Na_2S_2O_3$ times 0.3546

ClO_2 (chlorine dioxide)

(a) Put 30 mL of 10 percent KI plus 35 mL distilled water into erlenmeyer flask.

(b) Pipette 10 mL of dioxide sample into flask, being careful to hold pipette tip below liquid surface.

(c) Titrate with 0.1N $Na_2S_2O_2$ until colorless; record mL $Na_2S_2O_3$ as A.

(d) Add 10 mL 1:1 HCl (1 part concentrated hydrochloric acid to 1 part distilled water) and titrate to blue and continue titrating until colorless; record mL $Na_2S_2O_3$ as B.

(e) Grams per liter ClO_2 = (5/4A-1/4B) x 0.355.

(4) Figure grams wet pulp needed—divide grams o.d. pulp desired by percent solids.

(5) Figure total weight needed for desired consistency.

(6) Figure grams chemical needed.

(a) Multiply percent chemical times grams o.d. pulp (except when one chemical is sometimes based on grams of another chemical).

(b) Then divide grams needed by chemical strength in grams per liter to get mL needed.

(7) Figure makeup water needed—add grams wet pulp plus mL chemical and subtract from total weight needed = mL makeup water needed.

(8) Figuring chemical residuals

Cl_2 Residual

(a) Put 10 mL of 10 percent KI plus 10 mL of 10 percent acetic acid in erlenmeyer flask.

(b) Add 100 mL residual sample, titrate with 0.1N $Na_2S_2O_3$ until yellow, add starch to blue, and continue titrating until colorless.

(c) Grams per liter Cl_2 = mL $Na_2S_2O_3$ times 0.03546.

ClO_2 Residual

(a) Put 30 mL of 10 percent KI in erlenmeyer flask.

(b) Add 100 mL residual sample, plus 10 mL 1:1 HCl, titrate with 0.1 N $Na_2S_2O_3$ until yellow, add starch to blue and continue titrating until colorless.

(c) Grams per liter ClO_2 = mL $Na_2S_2O_3$ times 0.0169.

Chlorination (Sequential ClO_2/Cl_2 or D_C) Stage

(1) Both Cl_2 and ClO_2 are figured on o.d. pulp weight.

(2) Using a 10 L plastic bucket, add pulp, then add makeup distilled water at 20° F above desired reaction temperature. Start stirring motor and increase speed until best mixing is obtained.

(3) Add neeed ClO_2, making sure to keep good mixing. This work should be done under lab hood or in mill ventilated area; important to avoid breathing vapors.

94) After 2 minutes mixing add Cl_2 needed, again making sure good mixing is continued.

(5) Stop mixing after 10 min total time, take liquid residual sample and check final pH, Cl_2, and ClO_2 residual g per liter.

(6) Wash residual chemical from pulp; use warm water because this aids washing.

(7) Make brightness, viscosity, and K-No. pads.

(a) Dewater pulp in centrifuge and fluff pulp.

(b) Figure percent solids, then figure grams wet pulp needed to make 3 g o.d. buchner funnel pads. Weigh pulp needed for 3 pads.

(c) Fill blender ¾ full with polisher water, add pulp, blend on slow speed only long enough to slurry pulp.

(d) Start vacuum, put in filter paper, wet with distilled water, pour in pulp slurry.

(e) Start pad stack, 2 blotters, 1 metal plate. When most of the water is out of the pulp pad, remove from funnel and place on metal plate, filter paper up. Number as needed.

(f) Take one sheet, remove metal plate and dry for 1 hour on slow drier.

(g) Press 2 sheets at 1380 kPa for 2 min., then place in rings in humidity room. Place a larger filter over each sample sheet to keep it smooth and flat.

(h) Take sheet from slow drier after one hour and place in oven at 210° F for 20 min, then run viscosity and K-No. Check brightness after 24 hours drying. Figure percent residual by multiplying chemical residual strength in grams per liter times mL liquid in stage and dividing answer by grams of chemical added.

Oxygen Stage (O)

(1) Calculate o.d. grams of pulp remaining after above sampling by multiplying wet weight times percent solids.

(2) Figure total grams, grams NaOH needed and makeup water needed for desired consistency (usually 27 percent).

(3) Fluff pulp, place pulp in metal pan and spray with NaOH solution and makeup water in amounts necessary to get desired consistency and percent chemical on pulp evenly. (Usually 2 to 3 percent NaOH used.)

  (a) Spread pulp in pan and spray on NaOH and water mixture until evenly covered, stir pulp to expose dry pulp and spray again. Repeat until all NaOH and water is used.

  (b) Pound pulp with blunt instrument in such a manner that all pulp will be pressed to evenly distribute liquid to all pulp. Then run pulp through fluffer again.

(4) Preheat digester and baskets for around 3 minutes by direct steaming in digester. Carefully remove baskets, place pulp in baskets, approximately 75 o.d. grams in each basket, and place basket in digester. Place top on digester, tighten nuts, secure valve stack pipe onto rupture disc housing and start steam with relief valve and blow valve open. Steam slowly until temperature reaches 104° C.

(5) Close relief valve and blow valve, steaming slowly until temperature reaches 127° C. Stop steam and set O_2 pressure at 810 kPa in tank and connect O_2 line to digester with O_2 valve on digester closed. Open relief valve and blow valve to relieve pressure to 0. Close relief valve and blow valve and open O_2 valve. After O_2 pressure is reached maintain 121° to 127° C temperature.

CAUTION

If pressure begins to rise past 910 kPa with steam off, open blow valve and relieve pressure, shut off O_2. There may be a fire causing this and the aim is to avoid a pressure increase that would blow the rupture disc.

(6) After 30 min shut off steam and O_2. Let pressure off digester slowly with blow valve. Remove rupture valve stack pipe, remove digester top. With basket handle remove baskets. Dump and wash pulp (with warm water) into cheese cloth lined bucket. Run residual pH on 5 liter volume. Make brightness, viscosity, and K-No pads as before and after centrifuging and fluffing.

D-Stage

(1) Figure wet grams needed for desired o.d. pulp.

(2) Figure total grams needed for desired consistency.

(3) Figure grams ClO_2 by multiplying percent times o.d. pulp.

(4) Figure grams Cl_2 by multiplying percent times grams ClO_2.

(5) Figure makeup water needed.

(6) Plug in constant temperature bath and set at desired temperature.

(7) When temperature is up and level, place pulp in double plastic bags, add needed distilled water and mix well with pulp, then add needed chemicals together. Close off bag, remove air and tie top closed. Start time and place in bath, hold submerged with a brick.

(8) Near end of trial time place cheese cloth in bucket and get chemicals and glassware needed for pH and residual checks. At end of time dump pulp into cheese cloth lined bucket, squeeze residual into breaker, run pH and residuals.

(9) Wash pulp, centrifuge, fluff and make pads.

Pilot CEDED sequence bleaching procedure—General information and C-stage particulars are the same as for D/COD except:

(1) All information regarding the addition of ClO_2 can be ignored since only chlorine is added for the total reaction time.

(2) Chlorine is added at the beginning of the reaction time rather than after 2 min after beginning the stage since no chlorine dioxide is used.

E-Stage

(1) Calculate o.d. grams of pulp remaining after previous stage by multiplying wet weight times percent solids.

(2) Figure total grams, grams NaOH needed and makeup water needed for desired consistency (usually 12 percent).

(3) fluff pulp and place in double plastic bags and place in constant temperature bath.

(4) Add NaOH needed to makeup water and stir to mix. Use water at about 11° C above desired reaction temperature.

(5) Pour NaOH water mixture over pulp in double plastic bags and thoroughly mix by kneeding pulp through plastic bag.

(6) At the end of reaction time, wash pulp with warm water. Make brightness, viscosity and K-No pads as mentioned in D/C stage procedure.

D-stage—Same as last D-stage in D/COD procedure.

E_2-stage—Same as E_1-stage.

D_2-stage—Same as D_1-stage.

AIR SEPARATION PLANT

Oxygen purity was measured in an apparatus similar to an Orsat analysis kit. The oxygen purity measurement is based on the chemical reaction between the oxygen in a measured sample of gas and a highly activated form of copper contained in a water solution of ammonium hydroxide.

WASTEWATER TREATMENT PLANT

(1) Biochemical Oxygen Demand According to Standard Methods For The Examination of Water and Wastewater (Current Eddition at Time of Test)—Method 507. Note—Soluble BOD_5 is determined by filtering sample on Millipore AP40 filters and running BOD_5 on filtrate.

(2) Initial Settling Velocity—According to Standard Methods For The Examination of Water and Wastewater (Current Edition at Time of Test)—Method 213D.

(3) Sludge Volume Index—According to Standard Methods For the Examination of Water and Wastewater (Current Edition at Time of Test)—Method 213E.

(4) Mixed Liquor Suspended Solids and Volatile Suspended Solids (MLSS and MLVSS) and Total Suspended Solids and Volatile Suspended Solids (SS and VSS) According to Modified Gouch Crucible Technique NCASI Technical Bulletin No. 230, December 1969, "A Guide to the Determination of Biochemical Oxygen Demand Dissolved Oxygen, Suspended and Settleable Solids and Turbidity of Pulp and Paper Mill Effluents". This method used until March 1976. After March 1976 according to procedure outlined below. This procedure was developed from work done by NCASI as summarized in NCASI Technical Bulletin 291, "A Study of the Effect of Alternate Procedures on Effluent Suspended Solids Measurement", March 1977.

Preparation of Filter

(1) Label an aluminum weighing dish for each filter.

(2) Withdraw a Millipore AP 40 Microfiber Glass filter from the package. Note: At all times handle the filter and weighing dish with forceps.

(3) Center the filter on the support screen of a Sterifil filter holder. Screw the filter holder funnel firmly into place (be careful not to tear the filter).

(4) With vacuum applied to the filtration flask, wash the filter with 100 mL of deionized water.

(5) After all rinse water has been drawn through the filter, disconnect the vacuum.

(6) Transfer the filter (with forceps) to a numbered weighing dish and dry in an oven at 103 to 105° C for one hour.

(7) If you plan to determine volatile suspended solids, transfer the filter to a muffle furnace and ignite it at 550° C for 15 min. If not, omit this step.

(8) Cool filter and dish in desiccator.

(9) Weigh each filter and dish (combined) to the nearest 0.1 mg and record weight.

Determination of MLSS, MLVSS, SS and VSS

(1) Center the preweighed filter in the filter holder.

(2) Collect a sample (use a wide mouth pipet) of thoroughly mixed wastewater. Use a sample volume which will filter freely (no noticeable reduction in filtering rate).

(3) Pipet the sample in to the filter holder. (Vacuum must not be applied when adding the wastewater.)

(4) Apply vacuum and draw all of sample through the filter.

(5) Release the vacuum for one min.

(6) Add 10 mL of deionized rinse water to filter.

(7) Apply vacuum until all wash water is filtered.

(8) Release the vacuum for one min.

(9) Add 10 mL of rinse water. Apply vacuum.

(10) Release vacuum.

(11) Add 10 mL of rinse water.

(12) Apply vacuum until all rinse water is filtered.

(13) Release the vacuum and remove the filter funnel.

(14) Use smooth-tipped forceps to transfer the filter to its weighing dish.

(15) Dry the filter in an oven at 103 to 105°C for 1 hour.

(16) Cool filter and weighing dish for 30 min in a desiccator.

(17) Reweigh the filter and weighing dish (combined) and record its final weight.

(18) Subtract the initial weight from the final weight to determine the weight of collected residue. Report the results in mg/L using the following formula:

$$\text{TSS in mg/L} = \text{collected residue} \times 22.2222$$

$$= \frac{(\text{Wt. final} - \text{Wt. initial}) \times 1000}{45 \text{ mL}}$$

(19) For determining volatile suspended solids filter should be transferred to a muffle furnace and ignited for 15 min at 550°C. Then carry out steps 16, 17, and 18.

Color—

According to Tenative Procedure for Color Measurements of Pulping Wastes and their Receiving Waters—Spectophotometric Method. NCASI Technical Bulletin No. 253 December, 1971 "An Investigation of Improved Procedures for Measurement of Mill Effluent and Receiving Water Color".

Alkalinity—

According to Standard Methods for the Examination of Water and Wastewater (current edition at time of test)—Method 403 (pH 3.7 used an end point).

Sulfides for Hydrogen Peroxide Trials—
According to Standard Methods for the Examination of Water and Wastewater (15 edition). Method 428C using Hach Kit.

Dissolved Oxygen—
DO measured by a portable dissolved oxygen meter. Meter uses a probe which contains a semi-permeable membrane. The meter has an automatic temperature compensator. Also the sample is gently agitated to always pass a fresh sample of wastewater across the probe membrane. The DO analyzer is calibrated by the "Winkler" wet chemical method.

BLACK LIQUOR OXIDATION

Sodium Sulfide

Sodium sulfide was determined according to mecuric chloride potentiometric titration technique as outlined in NCASI Technical Bulletin No. 68, October, 1973.

Percent Black Liquor Solids

Percent black liquor solids was determined according to the following procedure:

(1) Duplicate samples are run on each set of evaporator discharges.

(2) Into a weighed ± 1 mg) aluminum dish is placed approximately 3 g weighed to the nearest 1 mg and recorded.

(3) The dishes are placed in the oven for 4 hours at 102°C (22°F).

(4) After 4 hours the aluminum dishes are placed in a desiccator to cool.

(5) The samples are then weighed to ± 1 mg and the percent solids calculated.

Calculations

$$\frac{\text{Dry weight of sample}}{\text{Wet weight of sample}} \times 100 = \text{percent solids}$$

APPENDIX C

DETAILED SYSTEMS DATA

This Appendix presents detailed data obtained on the systems listed below. Pertinent discussion is included. Where appropriate, pre- and post-grant period data and discussions are also presented.

(1) In-Process Changes

(2) Bleaching System

(3) Air Separation Plant

(4) Wastewater Treatment Plant

(5) Black Liquor Oxidation System

IN-PROCESS CHANGES

As indicated in earlier sections of this report, while Chesapeake was searching for the latest in wastewater treatment technology, decisions were made to expand into bleached hardwood market pulp. The timing was such that the bleach plant would be ready to start-up about one year before pilot trials were completed and a wastewater treatment plant could be engineered and installed. Accordingly, Chesapeake submitted (30) the overall plan to the SWCB which approved (6) the expansion, provided in-process BOD_5 reductions could be made to offset the projected, untreated bleach effluent. Though difficult to measure the total effect of such changes accurately, the following ones were made:

Replacement of Screw Presses With Vacuum Washers

This change was made to improve washing of batch digester pulp and, thereby, to reduce sodium and BOD_5 discharges from the pulp mill screen room and elsewhere. Chesapeake committed to the SWCB to reduce BOD_5 by 3859 kg/day (8500 lb/day) through washing system replacement. The study (31) to corroborate the BOD_5 reduction indicated 3,270 kg/day (5,000 lb/day) reduction at low hardwood production levels and 6,356 kg/day (14,000 lb/day) at high hardwood production levels. With the bleach plant operating, the batch system in the pulp mill would be in a high hardwood production mode. Therefore, the washer reduction commitment to the SWCB was more than met.

Installation of Steam Stripper

From measurements made over a period of ten years Chesapeake recognized that evaporator and blow heat condensate and turpentine decanter underflow streams were rich in BOD_5. To reduce waste load to the WWTP and odorous emissions at the WWTP cooling tower, Chesapeake explored both air and steam stripping. Since such a stripper would also be a vital part of the future odor abatement system an air stripper was ruled out because of the large air volume mixed with the stripped gases.

The stripper installed was designed to treat an average flow of 15.8 to 18.9 L/s (250 to 300 gpm) of turpentine decanter underflow and blow heat condensate (the surface condenser condensate was reused directly on the brown stock washers) with a BOD_5 load of 7264 kg/day (16,000 lb/day). The stripper was started up January 18, 1973 to treat only turpentine decanter underflow (5 to 5.6 L/s [80 to 90 gpm)] containing an average BOD_5 loading of 3950 kg/day (8700 lb/day). At 50 percent reduction of the BOD_5 by the stripper, the projected BOD_5 drop would be 1975 kg/day (4350 lb/day). Figure C—1 is a flow diagram of the stripper system.

An early study(32) of stripper performance showed 60 to 95 percent methanol (MeOH) removal (corresponds to BOD_5 removal) at 5 to 10 kg steam per 100 kg feed. Reduction in total reduced sulfur (TRS) and turpentine ranged from 80 to 99 percent. Additional data collected and reported(33) in September 1973 showed the following relationships between percent MeOH reduction, percent TRS reduction and reflux temperature, t_R, in the range 41 to 71°C (105 to 160°F).

$$\begin{aligned}\text{Percent MeOH reduction} &= 1.46\ (t_R,\ ^\circ\text{C}) - 47.51 \\ &= 0.812\ (t_R,\ ^\circ\text{F}) - 73.49\end{aligned} \qquad (1)$$

$$\begin{aligned}\text{Percent TRS reduction} &= 0.93\ (t_R,\ ^\circ\text{C}) + 21.59 \\ &= 0.514\ (t_R,\ ^\circ\text{F}) + 5.14\end{aligned} \qquad (2)$$

At $t_R = 71^\circ C$, percent MeOH reduction = 56 percent and percent TRS reduction = 88 percent.

During the September-1973 study a good correlation between methanol and BOD_5 concentration was established. This was expressed as follows:

$$BOD_5,\ \text{mg/L} = 1.195\ (\text{MeOH, mg/L}) + 448.8 \qquad (3)$$

The intercept of 448.8 is quite low compared with the MeOH and BOD_5 concentrations measured. At MeOH = O the remaining concentration of odorous compounds and turpentine accounted for the intercept of BOD_5. Figure C—2 pictures the BOD_5 versus MeOH correlation.

From equation (3), which is linear, it is evident that percent MeOH reduction and percent BOD_5 reduction are numerically equal. Thus, the 56 percent MeOH reduction found at a reflux temperature of 71°C in the mid-September-1973 study corresponds to a BOD_5 reduction, via stripping of turpentine decanter underlfow only, of (0.56) (3950) = 2212 kg/day (4872 lb/day). Although this is more than the projected 1975 kg/day (4350 lb/day), it is too low for one reason and maybe too high for another. First, the normal operating reflux temperature at the stripper was 93°C (200°F) or higher between bleach plant and WWTP startup. Second, the average pulp mill tonnage during the same period (1973 to 1974) was 961 to 979 metric tons/day (1060 to 1080 tons/day) instead of 1110 metric tons/day (1225 tons/day) assumed in the computation of projected BOD_5 drop across the scrubber. The higher reflux temperature means a percent MeOH (and percent BOD_5 reduction) much higher than 56 percent and the lower tonnage rate means less BOD_5 fed to the stripper. Adjusting the 3950 kg/day (8700 lb/day) BOD_5 at 1110 metric tons/day (1225 tons/day) to a production rate of 979 metric tons/day (1080 tons/day) gives 3482 kg BOD_5/day (7670 lb BOD_5/day). Assuming 80 percent reduction at 93°C (200°F) gives 2787 kg BOD_5/day (6140 lb BOD_5/day) reduction.

No further study was made of the stripper performance during the grant period due to the press of work being done on the oxygen bleach plant and in preparation for startup of the WWTP. However, it is reasonable to conclude from the data acquired that the steam stripper reduced BOD_5 to the extent expected.

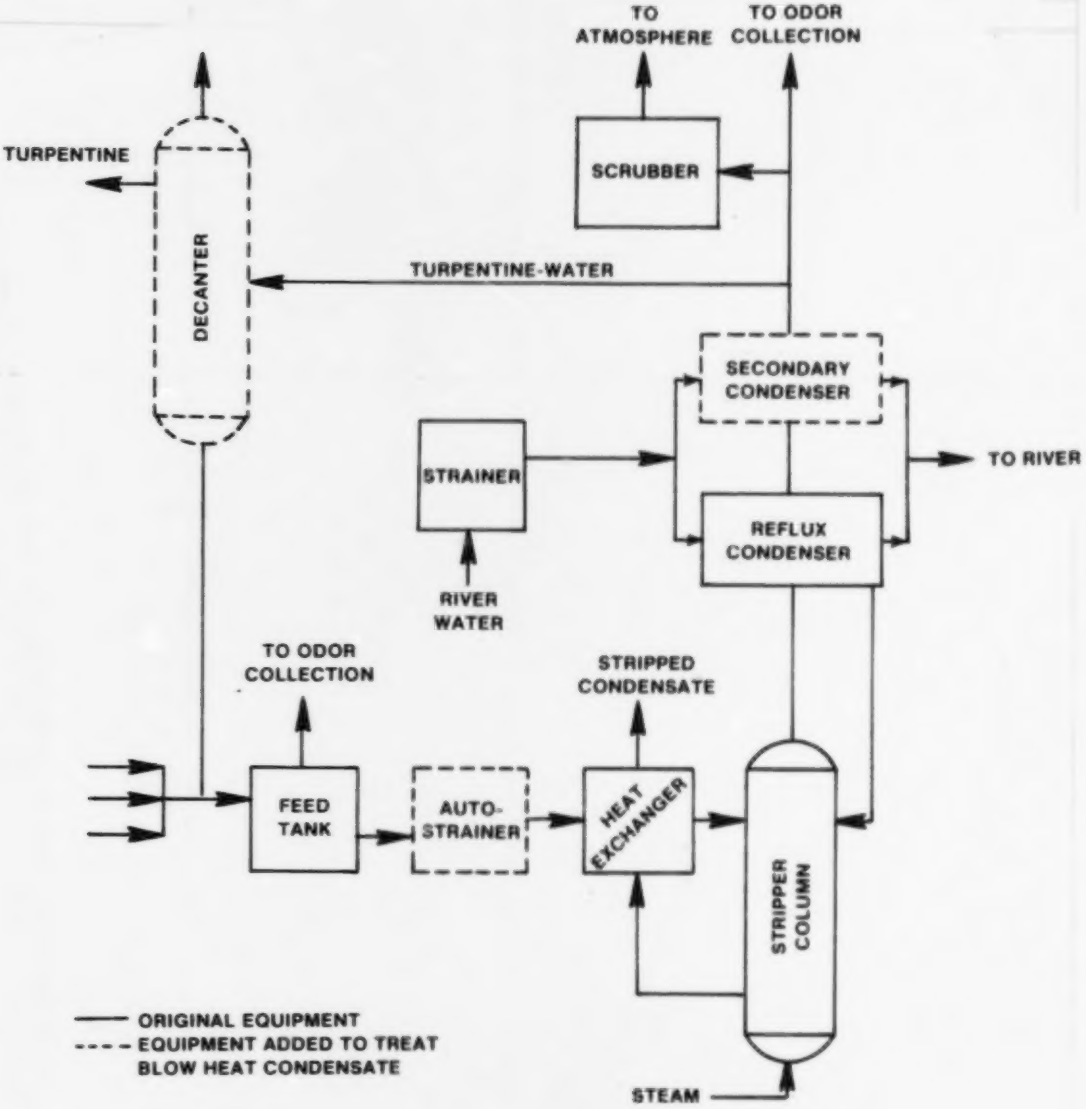


Figure C—1. Steam stripper flowsheet.

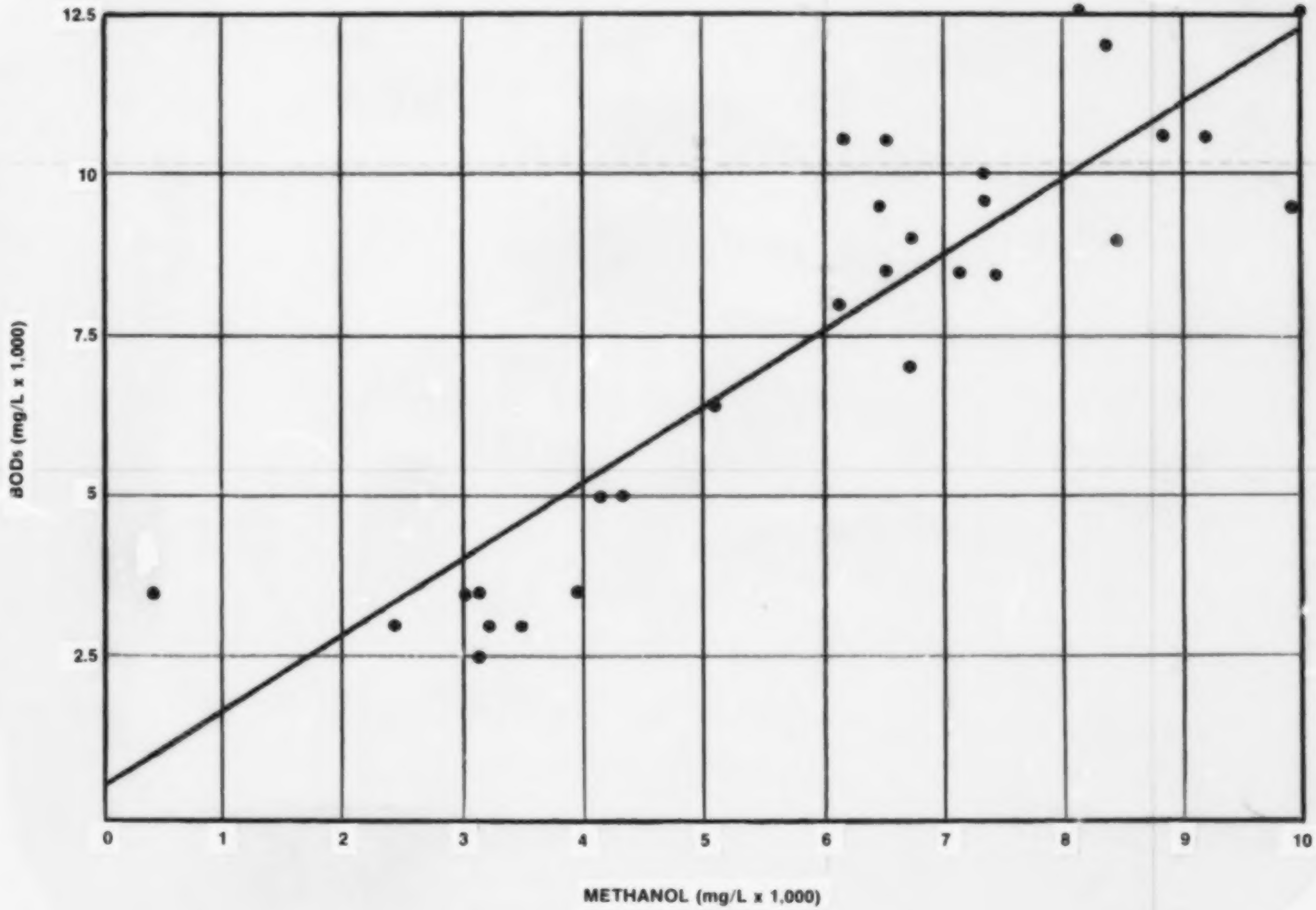


Figure C—2. Relationship between methanol and BOD$_5$.

As already mentioned, Chesapeake considered the steam stripper to be an integral part of the odor abatement system, which became operational in late January 1976. Obviously, the stripper started up before that date. Therefore, it was necessary to handle the off-gases in some manner until they could be incinerated. Two methods were considered, namely (a) piping to the tall recovery stack and (b) scrubbing the off-gases with white liquor or caustic solution. Method (b) was chosen.

Before the off-gases could be incinerated, it was necessary to condense their water vapor content. It was also desirable to collect any stripped turpentine. When blow heat condensate was added to the stripper feed, it was probable that small wood chips or fibers would have to be removed from the feed stream. Therefore, a secondary condenser, a turpentine decanter, and an automatic strainer were added to the stripper system to achieve the aforementioned objectives.

The principal mechanical problems in the stripper system occurred in the stripper feed heat exchanger shown in Figure C—1. This unit was plagued with gasket failures and tended to plug. The rubber gaskets in the original equipment were replaced several times then asbestos gaskets were substituted. After these failed at least twice, a spiral heat exchanger was installed to replace the original plate and coil exchanger. This solved both the plugging and gasket failure problems.

The secondary condenser was susceptible to plugging on the river water side, thereby diminishing its effectiveness.

Installation of Surface Condensers in the Evaporators

This in-process change from direct contact condensers was expected to reduce BOD_5 minimally although the net reduction would be extremely difficult to quantify as is indicated below. The main effect of the change was to reduce warm, brackish wastewater flow to the WWTP by 442 to 631 L/s (7 to 10,000 gpm), depending on the time of year.

Some reduction in overall BOD_5 discharge might occur if entrainment from the first evaporator effect could be totally eliminated by the new entrainment separators discussed in the next subsection. Reuse of the surface condenser condensate on the brown stock washers was certain to reduce BOD_5 to a degree not easy, if not impossible, to measure.

The surface condenser installation was completed December 28—29, 1972. By January 16, 1973 floating material from the river had passed through new salt water screens and plugged the condenser tubes. Showers were set up to wash down the screens and this step corrected the problem temporarily. The showered screens kept out floating matter but in March 1973 the 25.4 mm (1-in.) condenser tubes were being plugged again by submerged materials such as marsh grass roots from upriver. Plugging was also caused by river bottom mud stirred up by freighters and tugs berthing and leaving Chesapeake's shipping docks. Further plugging was reported(34) during the first year after installation. Some evidence of bacterial growth also appeared.

Shortly thereafter a river water chlorinator was installed to prevent "slime" growth in the tubes. Also, water lines were piped to enable back flushing of the surface condensers periodically. By chlorinating periodically, back flushing twice monthly, and cleaning three to four times yearly, the surface condensers have operated quite satisfactorily.

Installation of Entrainment Separators in the Evaporators

The opening comments of an internal report(35) on plans to investigate entrainment separators stated that the liquor carryover in the last effects of all three sets of evaporators contributed both color and BOD_5 to the waste discharge. In addition, this carryover was sufficient to limit the reuse of the last effect condensate that would be available for possible reuse if surface condensers were installed on each of our evaporators. In order to reduce the amount of carryover as a first step to enable reuse of last effect condensate, it was decided to install a demister in A-set evaporators on a trial basis. Experience gained as a result of this demister installation served to guide further installations.

A trial was undertaken and reported(35) on the last effect of one set of evaporators. The demister was placed in operation and performed satisfactorily until a section of the supporting structure (of the demister) fell to the bottom of the vapor head. Packing from the demister pad entered the liquor line leading to the next effect. This packing plugged the pump and caused the evaporator to be shutdown.

The following conclusions were drawn after discussions with operating people and the vendor, and study of the operating data before and after the demister structure gave way.

(1) Operating data indicate that plugging of the demister did not cause the pad to fall. Rather, it appeared that the supporting structure gave way as a result of failure of the pump pumping liquor to the fourth effect. When this pump failed, the demister was floated by liquor from the feed pump. When this liquor was pumped from the vapor head, the demister fell and its supporting structure broke.

(2) The continuous operation of a demister pad in A-set evaporators was questionable due to liquor foaming and associated collection of soap and fiber in the demister packing.

(3) Other means of reducing liquor carryover from A and C-sets evaporators appeared to be more promising as a long-term solution than a demister installation. These included:

   (a) Converting A and C-sets to sextuplet operation by adding new last effects when surface condensers were installed. (Surface condensers were needed to reduce flow to treatment plant by approximately 50 percent.)

   (b) Enlarging the vapor heads in the present last effects (5th) of A and C-sets when surface condensers were installed.

   (c) Installing more efficient external centrifugal separators when surface condensers were installed.

The cost and effectiveness of these alternate solutions was discussed with vendors before reaching a decision on the reinstallment of a demister in A-set.

A combination of recommended steps (3a), (3b), and (3c) was taken on evaporators sets A, B, and C when the surface condensers were installed. On A and C-sets, each of the old bodies of the direct contact condensers was converted to a centrifugal type entrainment separator after the last effect and before the surface condensers. On B-set, a parallel last effect was added with a larger vapor body to slow the velocity to the condensers. A new D-set of evaporators was equipped with a much larger vapor body to reduce the velocity to the surface condenser. Also centrifugal catch-alls which wrapped around the surface condensers were installed with the new surface condensers on A, B, C, and D sets of evaporators.

It is difficult to assign BOD$_5$ reduction values to each of the in-process changes described above. The combined effect can be seen in Table C—1, which lists the average daily BOD discharge for 19 four-week periods.

Table C—1 shows that, except for the period ending March 25, 1973, Chesapeake met the SWCB requirement of 19,068 kg BOD_5/day (42,000 lb/day). The March 25, 1973, period included a portion of the second startup of the bleach plant. During such a period, it could be expected that the untreated bleach plant effluent would be erratic and higher than normal. Therefore, it can be concluded that the in-process changes enabled Chesapeake to offset the expected bleach plant effluent of 4500 kg/day (10,000 lb/day).

Revisions in Pulp Mill Water System

As mentioned in Section 5 of this report, the bleach system explored involved overall countercurrent flow from pulp dryer back to the D-stage washer with internal recycles to reduce water requirements. Also having only two washers initially enabled lower water consumption. The goal was to install a bleach plant that would require no increase in the groundwater withdrawal rate.

The study of the pulp mill water system referred to in Section 5 resulted in changes that would provide the initial projected water requirements. Figures C—3 and C—4(36) show the general water system before and after bleach plant installation.

For comparison purposes, process water flow from the power plant (after use of water for turbine condensing) to the pulp mill was kept constant at 284 L/s (4500 gpm). However, 12.6 L/s (200 gpm) of fresh water (before turbine condensing) was picked up elsewhere and is shown as "Pomona well" on Figure C—4. The other principal changes were (a) storage was provided for the turbine condenser water in the "old diffuser tank" to provide adequate, stable flow of seal water to the pulp mill and bleach plant, (b) reduction of seal water to the pulp mill from 69.4 to 50.5 L/s (1100 to 800 gpm), (c) change from 37.8 L/s (600 gpm) to dry machine seals to 11.36 L/s (1800 gpm) for both dry machine and bleach plant for cooling and seal water, (d) recovering 75.7 L/s (1200 gpm) of the cooling and seal water from the dry machine to increase flow to batch and Kamyr turpentine condensers from 44.2 and

TABLE C—1. AVERAGE DAILY BOD_5 DISCHARGE BETWEEN DECEMBER 4, 1972 AND MAY 19, 1974

28-day period ending	Average BOD_5 kg/day (lb/day)
December 31, 1972	17,463 (38,500)
January 28, 1973	16,268 (35,865)
February 25, 1973	14,733 (32,480)
March 25, 1973	19.840 (43,740)
April 22, 1973	16,057 (35,400)
May 20, 1973	17,300 (38,140)
June 17, 1973	16,918 (37,299)
July 15, 1973	18,519 (40,828)
August 12, 1973	16,652 (36,712)
September 9, 1973	12,895 (28,429)
October 7, 1973	14,984 (33,035)
November 4, 1973	17,817 (39,279)
December 2, 1973	14,518 (32,007)
December 30, 1973	16,932 (37,329)
January 27, 1974	11,404 (25,141)
February 24, 1974	15,660 (34,524)
March 24, 1974	17,380 (38,317)
April 21, 1974	16,970 (37,413)
May 19, 1974	16,604 (36,606)

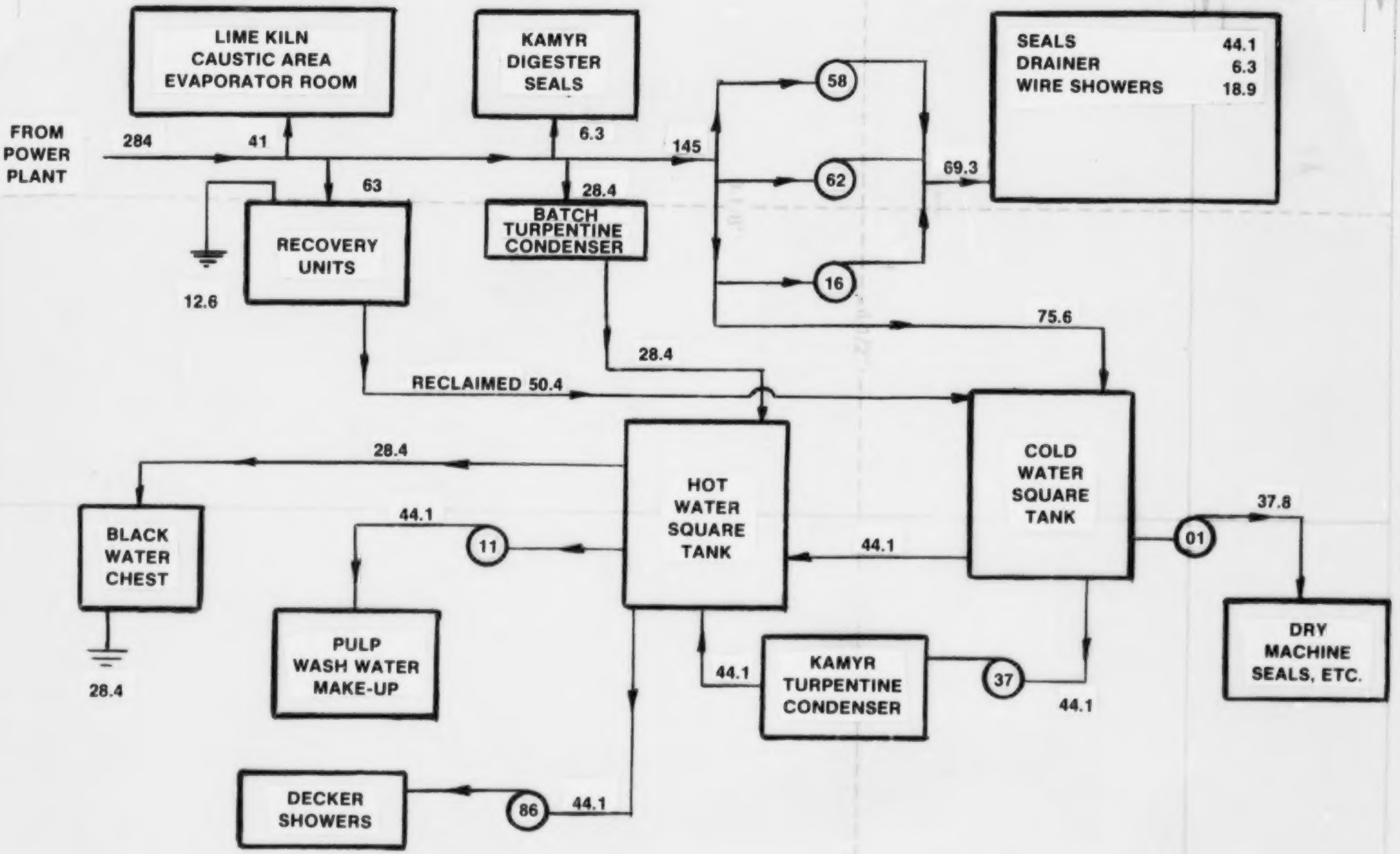


Figure C—3. Pulp mill process water system before bleach plant installation (all flows in L/s).

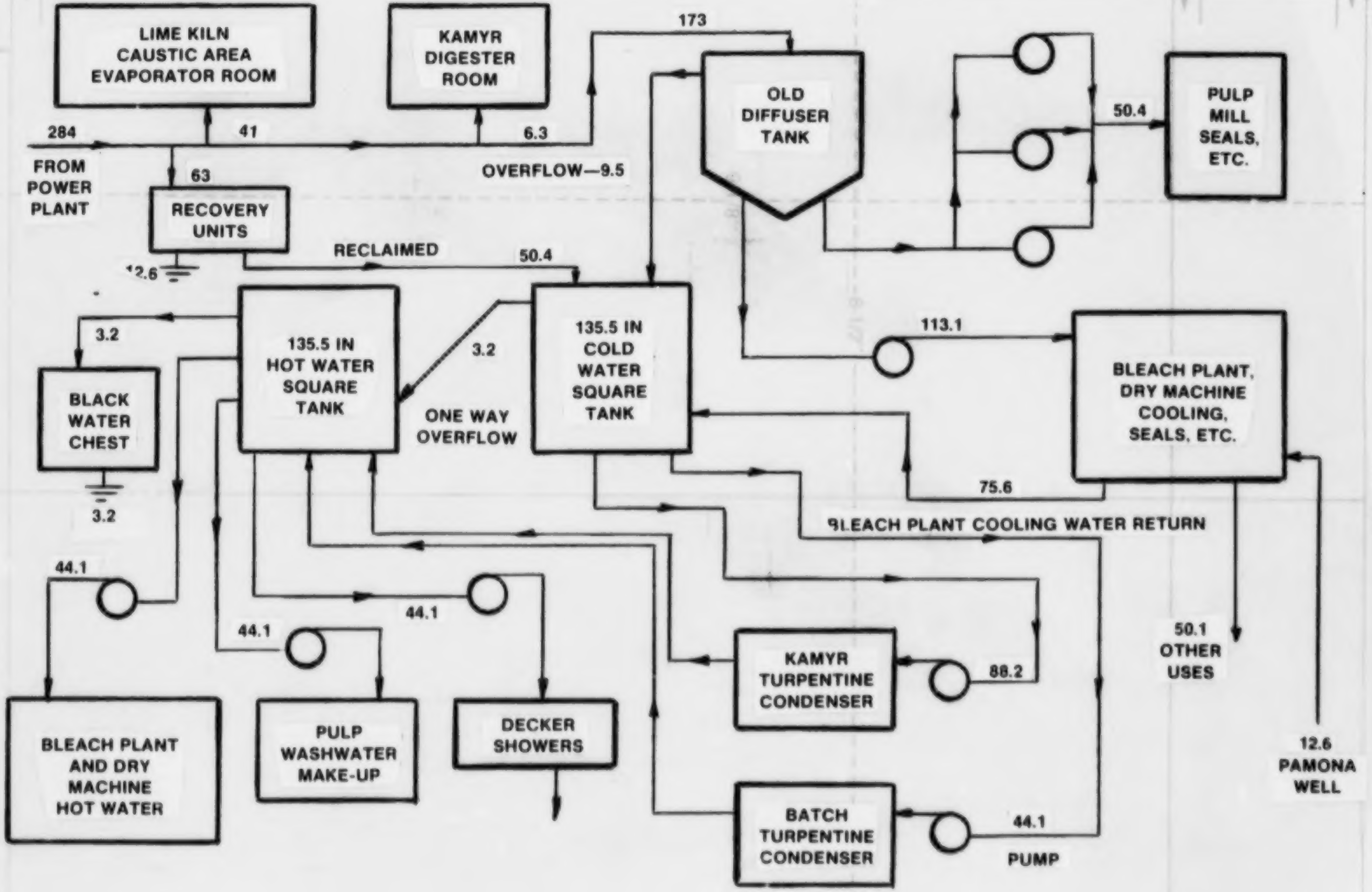


Figure C—4. Pulp mill process water system after bleach plant installation (all flows in L/s).

28.4 to 88.3 and 44.1 L/s (700 and 450 to 1400 and 700 gpm), respectively, and (e) providing 44.2 L/s (700 gpm) of hot water to the dry machine and bleach plant.

The net effect of these changes was to make 94.7 L/s (1500 gpm) available to the dry machine and bleach plant. This was sufficient to provide the initial estimated requirement of 97 to 104 L/metric a.d. ton (7000 to 7500 gal/a.d. ton) at 272 metric a.d. tons/day (300 a.d. tons/day) production.

After installation of the D/C washer, water requirements increased by 15.8 to 18.9 L/s (250 to 300 gpm). This quantity was also picked up elsewhere without increasing groundwater withdrawal rate.

BLEACHING SYSTEM

Sequence Selection

Chesapeake established the minimum specifications for its proposed bleached hardwood pulp to be 88+ GE brightness (essentially) numerically equal to Elrepho brightness and 14.0 centipoise minimum viscosity. Unbleached hardwood pulps submitted to L'Air Liquide, Paris, France for AO bleaching covered a range of permanganate numbers (K-Nos.*) from 10 (sampled from mill deckers) to 30. Samples at 20 and 30 K-No. were "hard-cooked" in a Chesapeake pilot digester to establish whether or not the AODED sequence would disintegrate and bleach exceptionally high shive and dirt content in the unbleached pulp.

Pilot digester unbleached pulp with 15.8 K-No. was sent to Canadian Industries, Ltd., McMasterville, Quebec for evaluation of the D/COD and other bleaching sequences. Table C—2 shows the pulp composition and some characteristics of the unbleached pulp submitted to CIL.

Note from Table C—2 that the unbleached viscosity of 36.5 cps was unusually high for a run-of-the-mill hardwood mixture. As with the sample sent to L'Air Liquide, the pulp sent to CIL was intended

TABLE C—2. NATURE OF UNBLEACHED HARDWOOD PULP SUBMITTED TO CIL

Type of unbleached pulp samples	Description	
Hardwood	Composition:	8% southern pine 92% hardwood
	Hardwood:	35 to 45% mixed oaks 55 to 65% sweet gum and yellow poplar
	Kappa No.:	21.7
	Viscosity:	36.5 cps
Softwood	Composition:	southern pine
	Kappa No.:	28.8
	Viscosity:	34.3 cps

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*The permanganate number (K-No.) is a measure of the residual lignin in the pulp after cooking the wood; the analytical procedure used to determine K-No. is included in Appendix B. Frequently, Kappa No. (See Table C—2) is also used as a measure of residual lignin. The Kappa No. vs. K-No. relationship is given by:

$$\text{Log (Kappa No.)} = 0.837 + 0.0323\ \text{(K-No.)} \qquad (4)$$

to study the effect of D/COD and other oxygen bleaching sequences on reducing shive and dirt content.

As indicated in Section 5, Chesapeake explored several conventional chlorination sequences. However, the tables of data below will show only the CEDED sequence compared with AODED and D/COD, the principal oxygen sequences studied.

Comparisons(37) made by Chesapeake on certain pine pulps prior to the studies by L'Air Liquide and CIL were informative and helpful in setting the course of further study. Table C—3 shows data on pine pulp sampled at Enstra, S.A. before and after the O-stage in the AODED sequence during Chesapeake's visit to the S.A.P.P.I. mill in January, 1971.

The basic strength properties were not significantly different before and after AO bleaching even at the abnormally low unbleached viscosity of 21.9. A viscosity of 25 to 35 was considered to be a more normal range for pine pulp.

Table C—4 shows similar data(37) on pine pulp sampled at Husum, Sweden before and after the O-stage during Chesapeake's visit to Mo Domsjo A.B. (MoDo) in March 1971. This visit was made after conferring with and hearing papers (38)(39)(40) presented by MoDo and CIL technical people at the Tappi National Conference in February 1971.

TABLE C-3. DATA ON S.A.P.P.I. PINE PULP BEFORE AND AFTER AO BLEACHING

Pulp property*	Before (unbleached)	After (AO bleached)
Breaking length, meters	10,287	9,779
Tear factor	116	108
Burst factor	87	89
Viscosity (cps)	21.9	11.7

*At 300 cc Canadian Standard Freeness (CSF)

TABLE C—4. DATA ON MODO PINE PULP BEFORE AND AFTER O_2-STAGE

Pulp property*	Before (unbleached)	After (O_2-bleached)
Breaking length, meters	11,174	10,992
Tear factor	83	75
Burst factor	104	107
Viscosity (cps)	42.0	24.5

*At 300 cc CSF.

Again, the differences in basic pulp properties were about the same before and after an O-stage bleach. The tear was slightly lower but the burst was higher. Viscosity of both unbleached and O-stage bleached pulps were higher than corresponding S.A.P.P.I. pulp viscosity. Greater emphasis appeared to be placed on higher viscosity by MoDo to preserve unbleached strength properties to the maximum extent possible in the bleached pulp.

In these two instances the unbleached viscosity dropped by 46.7 percent in the AO stage (S.A.P.P.I.) and 41.7 percent in the O-stage (MoDo). This difference may not appear significant based on these data alone but results presented below show the importance of high unbleached viscosity to attaining high bleached viscosity.

Table C—5 shows a comparison of pulp properties for a low and high K-No. hardwood pulp bleached by the AODED and CEDED sequences. Low K-No. pulp reported in Table C-5 was sampled from the Chesapeake pulp mill; the high K-No. pulp was prepared in Chesapeake's laboratory pilot digester. As mentioned above both pulps were sent to L'Air Liquide for AO bleaching only. Chesapeake completed the DED stages for the AODED sequence and carried out the CEDED sequence on the 10.5 K-No. pulp. Other sequences were explored by L'Air Liquide by the data reported in Table C—5 for AODED was the best and most representative.

Clearly from Table C—5, bleached viscosity was higher when unbleached viscosity (19.5 K-No.) was higher even though the overall percent drop, in this case, was 72 versus a drop of 60 percent for the lower (10.5 K-No.) unbleached viscosity pulp. With the CEDED sequence the unbleached viscosity drop was 64 percent overall to 7.5 cps. There appeared to be no advantage or disadvantage of AODED over CEDED, or vice versa, with respect to pulp quality.

Table C—6 shows Chesapeake data on the 15.8 K-No. pulp sent to CIL for evaluation of various oxygen bleach sequences. CIL completed the D/COD and CEDED sequences, and others indicated below, in their laboratories. Chesapeake as with the data in Tables C—3, C—4, and C—5, tested the bleached pulps for properties reported in Table C—6.

TABLE C—5. COMPARISON OF PROPERTIES OF LOW AND HIGH K-NO. HARDWOOD PULPS

Pulp properties*	Low (10.5 K-No.) pulp+			High (19.5 K-No) pulp‡	
	Unbleached	AODED	CEDED	Unbleached	AODED
Breaking length, meters	6,858	7,620	7,493	9,017	8,890
Tear factor	90	83	86	87	81
Burst factor	55	57	52	64	70
Viscosity, cps	20.6	8.3£	7.5	40.5	11.4#
Brightness, percent GE	——	90.4	90.2	——	90.3

*At 300 cc CSF

+ Pulp sampled from regular production in the pulp mill.
‡ Pulp prepared in laboratory digester.
£ AO viscosity was 9.9 cps.
# AO viscosity was 15.0 cps.

Again the data in Table C—6 show little difference in basic pulp properties between CEDED and D/COD. However, the viscosity differences are notable. The D/COD sequence gave a bleached viscosity of 16.5 cps versus a viscosity of 12.0 cps from the CEDED sequence at the same brightness and breaking length levels. Referring to Tables C—5 and C—6 it could be expected, because of the higher K-No. that the 19.5 K-No. unbleached pulp would have the higher bleached breaking length of 8890 meters compared with 8128 meters for the 15.8 K-No. pulp, so long as the bleaching sequences were reasonably similar. Despite the higher breaking length, however, the AODED sequence resulted in a much lower viscosity of 11.4 cps versus 16.5 cps for the D/COD sequence. Chesapeake concluded that the D/COD sequence would produce bleached hardwood pulp with higher viscosity than the AODED sequence at equal breaking length.

CIL also carried out studies on the COD, OD/CED and D/CODED sequences.(15) The bleached pulps resulting from the COD sequence were of slightly inferior quality to those produced by (D-C)OD. The sequence O(D-C)ED can be used to produce pulp of quality beyond the minimum specifications. However, within the scope of evaluation of the process conditions, it was impossible to produce the quality of pulp produced by (D-C)OD. On the other hand, the four-stage process was advantageous as to chemical cost and pollution abatement.

CIL carried out a second series of studies(41) on the D/COD and D/CODED sequences on the 15.8 K-No. hardwood. Softwood bleached by the D/COD sequence was also studied. In this second series, application of magnesium complex in the second stage (O-stage of D/COD) preserved viscosity significantly and resulted in an increase of 3 viscosity units for a fixed brightness (or over 2 brightness units for a fixed viscosity), as illustrated in Figure C—5. The process parameters in the O-stage were 27 percent consistency; 120°C (248°F); 620 kPa gauge (90 psig) O_2 pressure and 0.1 percent magnesium complex on oven-dry pulp expressed as Mg^{++}.

The effect of total available chlorine (expressed as a constant multiplied by the Kappa No., K.) and percent replacement by chlorine dioxide (on an available chlorine basis) on the brightness-viscosity relationship after the O-stage is shown in Figure C—6. One mole of Cl_2 and one mole of ClO_2 provides 1.0 and 2.62 moles of available chlorine, respectively. CIL reported that levels of 0.22 Kappa (0.22K) and 50 percent replacement, and 0.27K and 50 and 60 percent replacement yielded maximum second stage (O-stage) brightness over a range of acceptable viscosities. But 0.22K and 50 percent replacement was selected as optimal because of the obvious cost advantage.

The second CIL study(41) reported that pH after the second stage (O-stage) correlated linearly with brightness, providing a convenient and accurate control parameter for the second stage. This

Table C—6. CHESAPEAKE DATA ON HARDWOOD PULP (15.8 K-NO.) SENT TO CIL

Pulp properties*	Unbleached	CEDED	D/COD
Breaking length (m)	8,130	8,383	8,128
Tear factor	88	85	84
Burst factor	60	65	64
Viscosity, cps	36.5	12.0	16.5
Brightness, percent GE	—	90.4	90.3

*At 300 CSF.

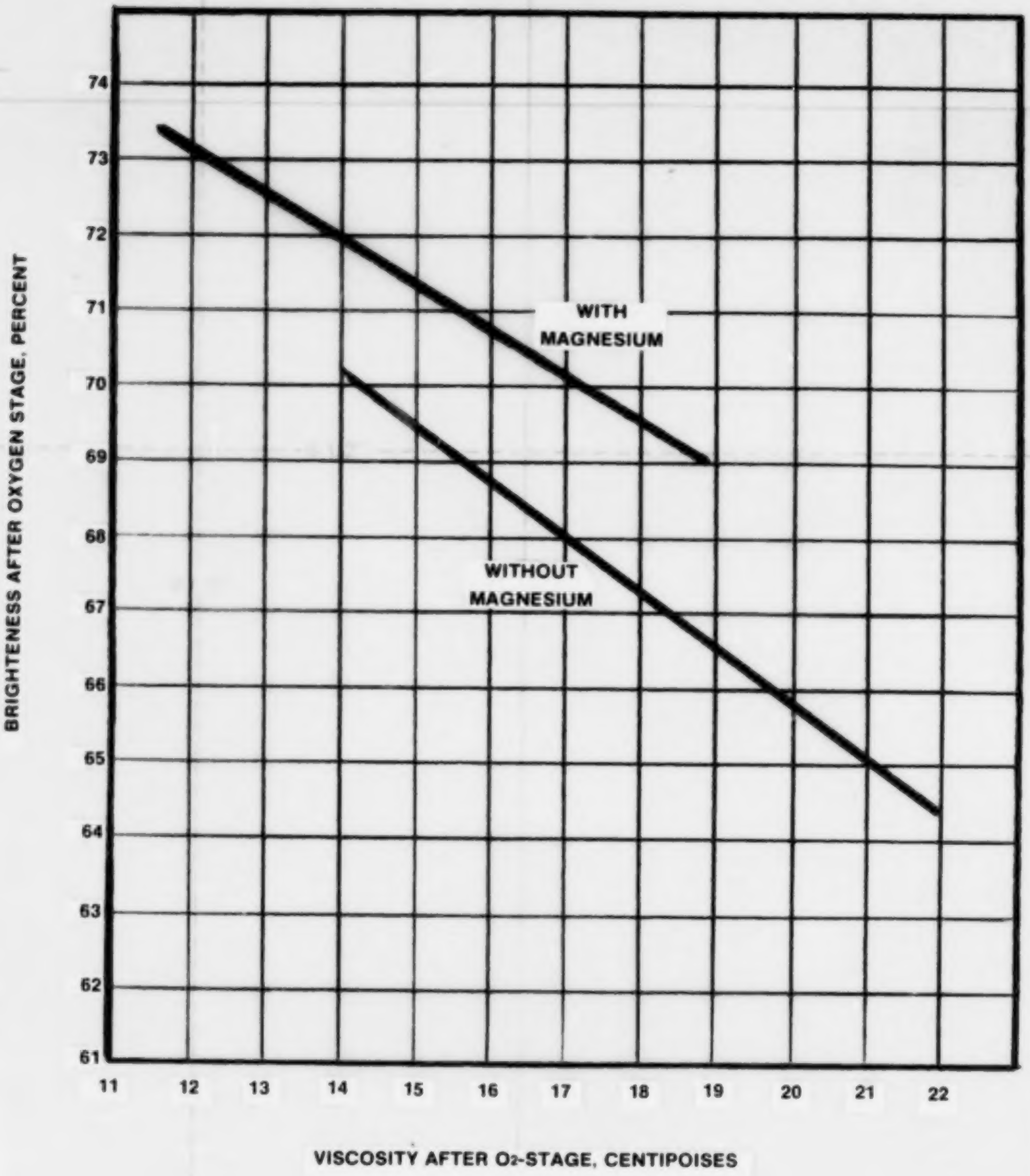

Figure C—5. Effect of magnesium on oxygen stage viscosity.

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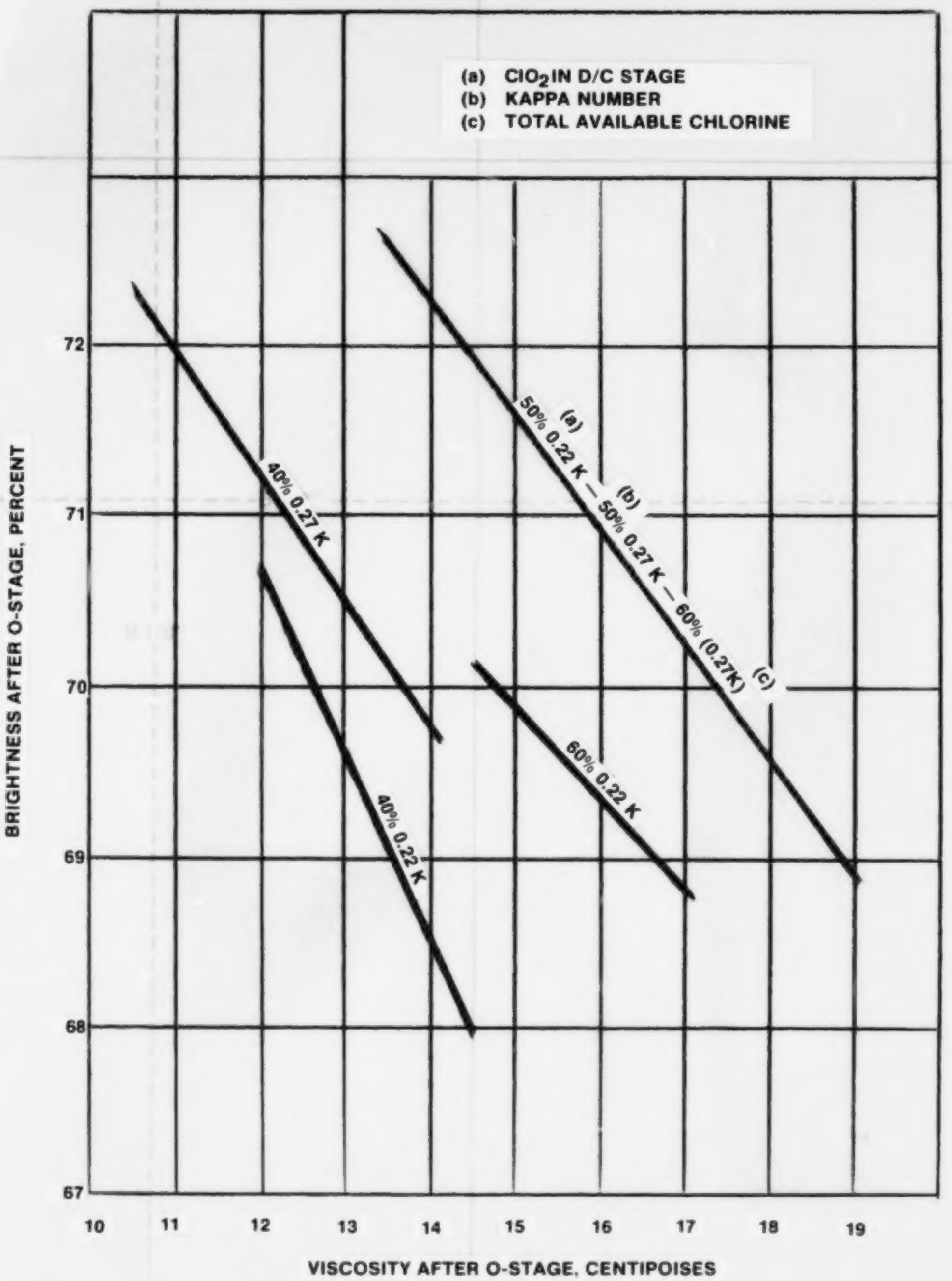

Figure C—6. Effect of total available chlorine and chlorine dioxide in D/C stage.(9)

relationship, observed both in the pilot plant at MoDo and in the laboratory at CIL, is shown in Figure C—7.

Considerable flexibility for the D/COD sequence was displayed in the CIL second study. Second-stage pulp quality of 70.0 brightness and 18 cps viscosity allowed a consistent 90+ brightness and 16 to 18 cps viscosity (after the third, or D-stage). A higher degree of bleaching (71+ and 15 cps) resulted in 91+ final stage brightness at target viscosities (14+) if a higher temperature (85° C) was used in the stage.

CIL's second report(41) on the study of the D/CODED sequence states that the extra two stages offered little or no advantage over the three-stage sequence with respect to maximum brightness, since the small chemical cost saving would not compensate for the increased capital cost for the two stages.

Regarding softwood bleaching with the D/COD sequence, the report(41) states that the softwood displayed an excellent bleach response to the three-stage sequence. With the use of retention times and temperatures established in the hardwood studies, brightness approaching 92.0 with nearly 19 cps viscosity was achieved.

From the standpoint of selecting a sequence to achieve the quality specifications of 88+ brightnesscps viscosity for bleached hardwood pulp, Chesapeake concluded that all evidence indicated D/COD as the best sequence.

With respect to pollution abatement, CIL's report(41) states that the three-stage sequence, when compared to a conventional five-stage process, has certain pollution abatement advantages. The total color in the three effluents is less than half the 90-kg/metric a.d. ton (180 lb/a.d. ton) of color expected from a five-stage sequence. Though the total BOD_5 is approximately the same in both three- and five-stage sequences, the high concentration in the oxygen stage necessitates only treatment (or recycle to chemical recovery) of this low volume effluent [4506 L/metric a.d. ton (1080 gal/a.d. ton)] to achieve a 60-percent reduction in BOD_5. Using less chlorine in the first stage decreases the effluent chloride, particularly in the oxygen stage, which can be compared to a conventional extraction stage. The latter would be expected to contribute more than twice as much chloride—44 kg/metric ton (88 lb/ton).

Although the D/COD sequence appeared to offer certain pulp quality advantages, such as bleached viscosity levels, the pollution abatement potential was not as great as the AODED sequence. Therefore, both sequences were seriously considered before final selection. Since neither system had been compared as to capital cost with the CEDED sequence, Chesapeake requested capital cost quotations on all three sequences. The relative estimated capital costs, taking the D/COD sequence as unity, were as follows:

D/COD	1
CEDED	1.2
AODED	1.4

Obviously, the D/COD sequence offered substantial first cost advantages. Although chemical costs were projected to be somewhat higher from the laboratory work, maintenance costs could be projected to be lower for a 3-stage versus a 5-stage system. Also, although AODED offered somewhat greater pollution abatement potential than the D/COD, the latter appeared promising from this standpoint, especially if the chloride level in the O-stage filtrate would be low enough to permit recycle to the mill chemical recovery system. Since the CIL and Chesapeake laboratory work did not simulate interstage pressing and washing the pollution abatement potential of the D/COD could be

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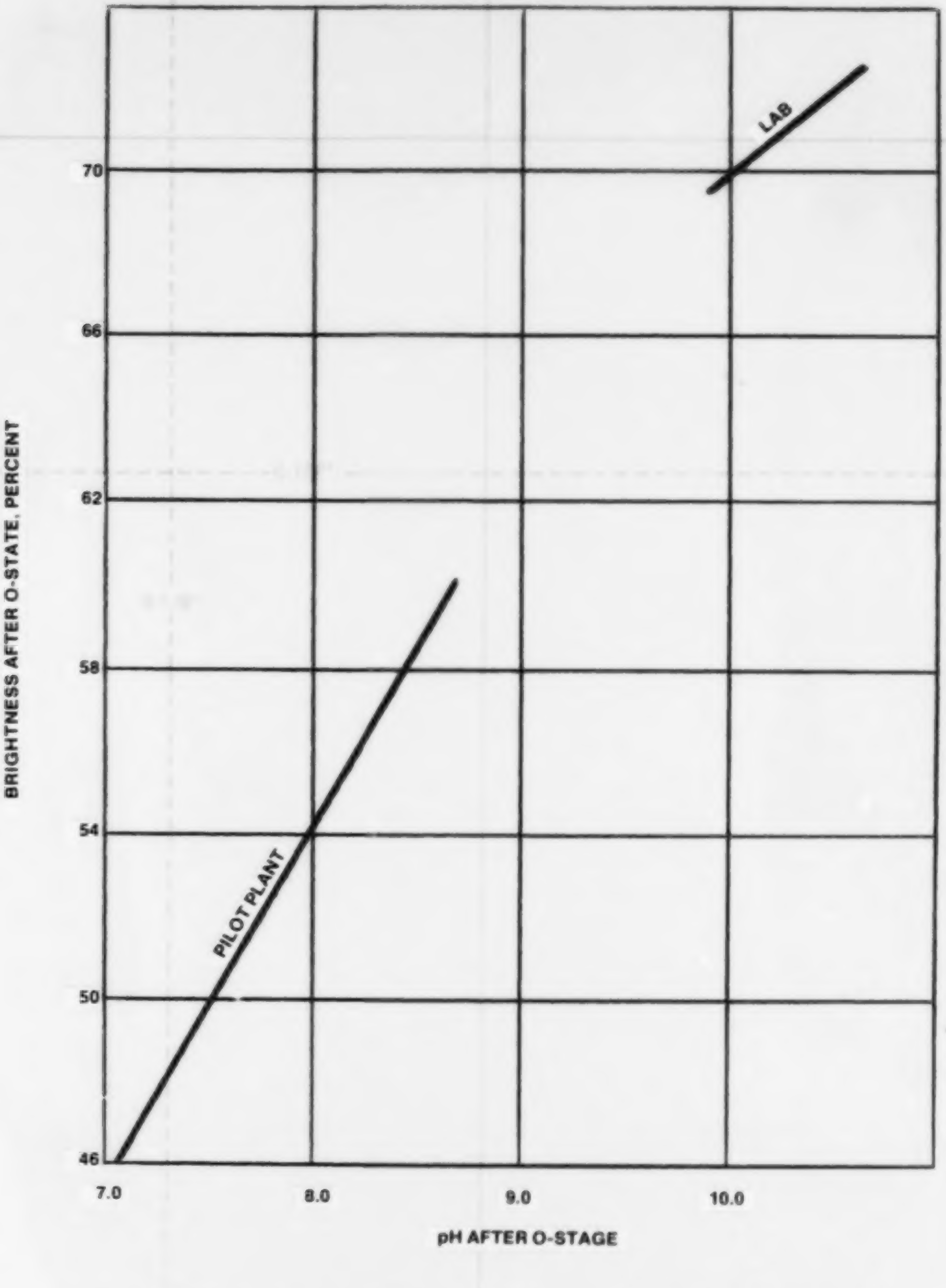

Figure C—7. Correlation of brightness and pH after O-stage. (41)

determined only from operation of a full-scale system. Should the chloride level be too high to permit O-stage filtrate recycle Chesapeake felt it had an option to remove chlorides from white liquor in the mill by a system of its own design or by a system such as the proposed Rapson-Reeves crystallization process(42). Another option available was to treat the total combined bleach effluent in an external waste treatment system.

Considering all cost and process factors Chesapeake chose the D/COD sequence to produce 272 metric a.d. tons/day (300 a.d. tons/day) average of bleached market hardwood pulp of 88+ brightness and 14 cps viscosity where the latter was required.

Process Design, Engineering and Installation

In the second study done at CIL(41), the process parameters for the sequence D/COD were optimized. Table C—7 lists the process operating conditions recommended by CIL for all three stages.

Certain of the operating conditions in Table C—7 are still practiced. However, several have been changed for a variety of reasons, as explained below.

At the time of design the unit costs of the basic chemicals used in bleaching were as shown in Table C—8.

TABLE C—7. RECOMMENDED OPERATING CONDITIONS FOR D/COD SEQUENCE*

Operating Parameter	Stage		
	D/C	O	D
Temperature, °C	50	120	80
Consistency, percent	3	27	12
Chlorine dioxide, percent	0.91	—	0.9
Chlorine, percent	2.39	—	—
Chlorine, percent	2.39	—	—
Caustic soda, percent	—	3	—
Oxygen, percent	—	1.5	—
Available chlorine, percent	4.77	—	2.36
Percent replacement with ClO_2	50	—	—
Residual chlorine, g/L	0.08	—	0.06
pH (out)	—	10.6	—
Retention time, min	10	30	300
Brightness, out	—	70	88+
Mg++ addition, percent	—	0.1	—
Pressure	Atm.	90 psi	

*Production rate average is 300 a.d. tons/day; peak is 330.

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With the data provided in Table C—7, CIL Ltd. engineered the D/COD sequence with input from Impco on the mixing, pressing, pumping and washing systems. A detailed description of the bleach plant has been published.(44) Figure C—8 pictures the flow diagram of the bleach plant as initially installed and Figure C—9 shows the plant as it is laid out presently (March 1978). The bleach plant was designed for an average production rate of 272 metric a.d. tons/day (300 a.d. tons/day) with a peak rate at 299 metric a.d. tons/day (300 a.d. tons/day).

The SVP chlorine dioxide generator selected by Chesapeake has been described by Hooker Chemical in the literature(43). The system for Chesapeake was designed to produce 6.8 metric tons/day of ClO_2 (7.5 tons/day), with a peak, short-term rate of 9.1 metric tons/day (10.0 tons/day).

Materials of construction were chosen to withstand typical and special corrosive situations in oxygen bleaching system. For example, the oxygen reactor was built of mild steel but lined with Carpenter 20 alloy. The first stage consisted of two mild steel tanks lined with fiber reinforced plastic (FRP). The third stage (D-tower, both updraft and downdraft) was fabricated of FRP alone. The oxygen washer was built of 316 stainless.

TABLE C—8. UNIT COSTS OF BLEACHING CHEMICALS (1971)

Bleaching chemical	Unit cost ¢/kg	(¢/lb)
Chlorine	1.8	(3.9)
Sodium hydroxide	1.9	(4.1)
Sodium chlorate	3.4	(7.5)
Sulfur dioxide	1.9	(4.2)
Oxygen	0.9	(2.0)
Sulfuric acid	0.5	(1.1)
Sodium chloride	0.5	(1.0)

Using these unit costs, the chemical consumptions projected in Table C—7 and the use factors published(43) for the SVP generator, an estimated bleaching chemical cost was calculated and is shown in Table C—9.

TABLE C—9. ESTIMATED COST OF BLEACHING CHEMICALS FOR THE D/COD SEQUENCE

Bleaching chemical	Cost per ton, $ Metric ton	(2000 lb/ton)
Chorine	1.69	(1.86)
Sodium hydroxide	2.23	(2.46)
Oxygen	.54	(0.60)
Chlorine dioxide	3 .58	(3.95(1)
Total	8.05	(8.87)(2)

(1) The estimated cost of chlorine dioxide does not include labor, taxes depreciation, etc. It does include the cost of sodium chlorate at a chlorate efficiency of 94.5 percent, or 1.66 kg of sodium chlorate per kg of ClO_2 produced. The cost also includes utilities cost as published(43 at 2.97 ¢/kg Cl_2 produced.

(2) Cost of chlorine dioxide under (1) was a net of 4.95 ¢/kg (10.91 ¢/lb). If the CIL estimate of 8.12 ¢/kg (17.9 ¢/lb) is used the total bleaching chemical costs become $11.40.

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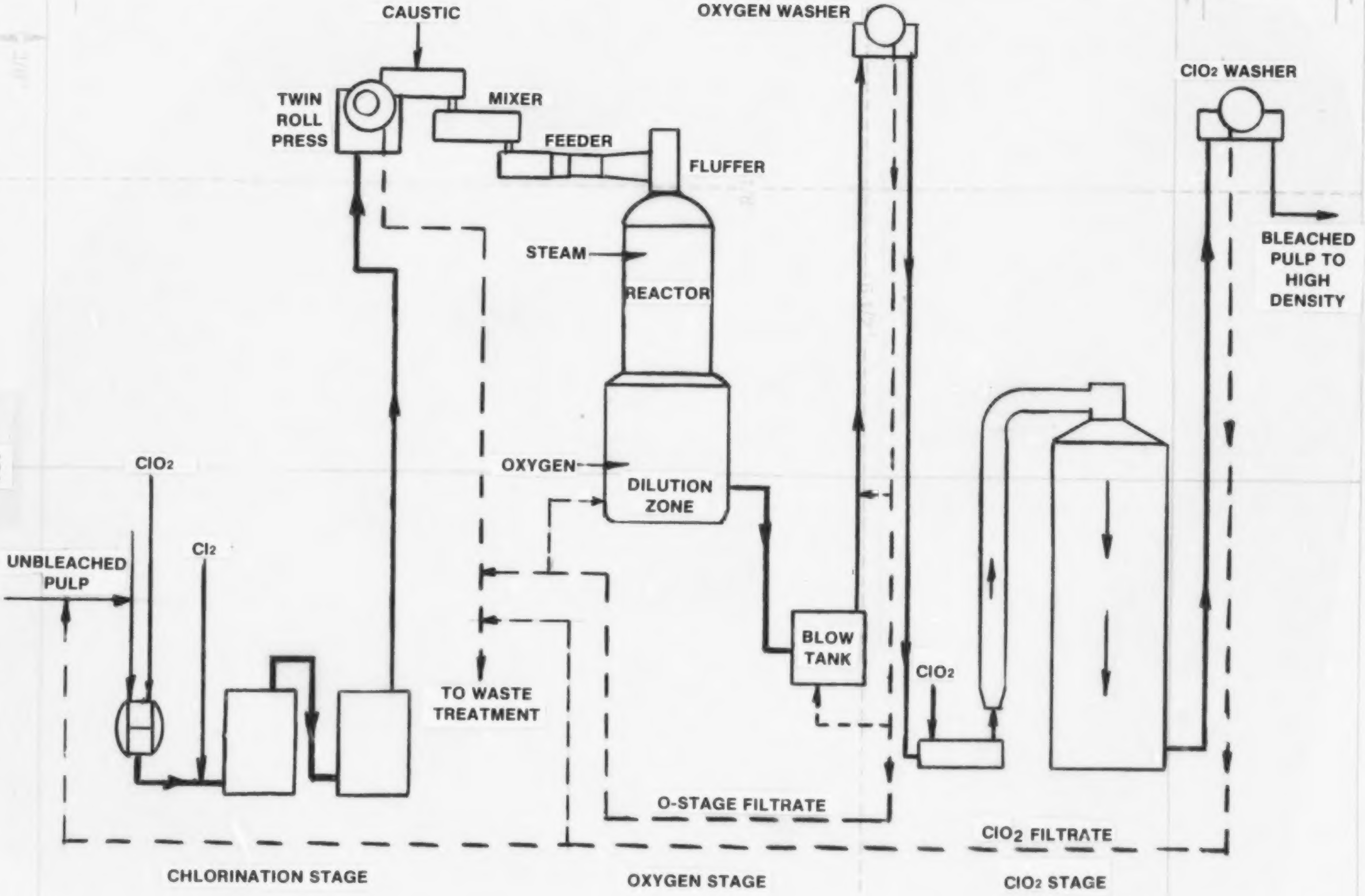

Figure C—8. Initial bleach plant flow diagram.

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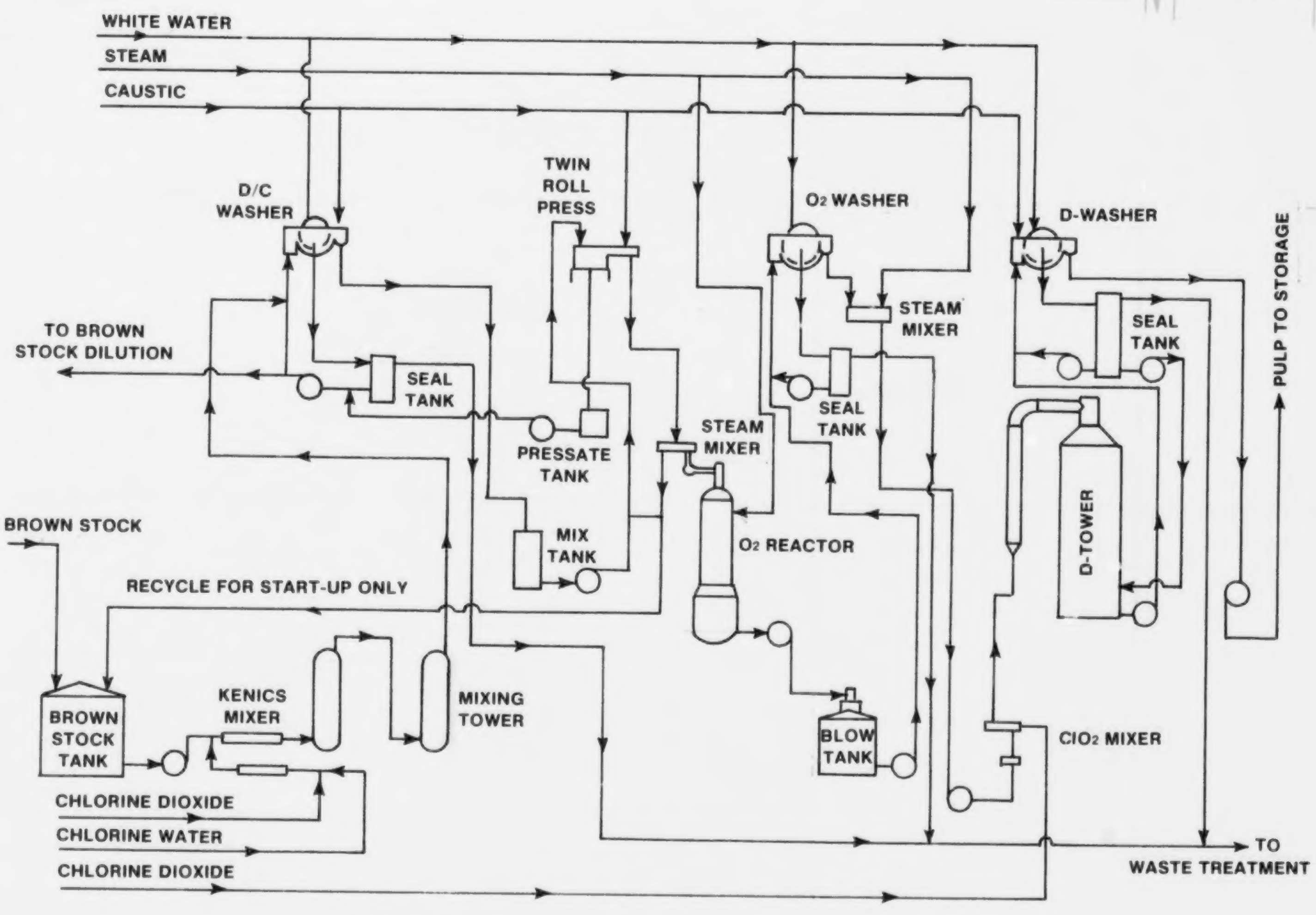

Figure C—9. Bleach plant flow diagram March 1978.
(Excluding second twin roll press.)

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Bleach Plant Evaluation

For sometime after both the initial startup (December 1972) and the second startup (late February 1973) evaluation of the bleach plant fell into four categories. These were (a) resolution of mechanical and process problems, (b) measurement and improvement in pulp quality, (c) study of process effluents by stage and combined, and (d) assessing the possibility of recycling the O-stage filtrate to the mill recovery system. Results and discussions related to these categories follow. Some data are presented from the post-grant period especially when such data lead to conclusions that differ from those reached during the grant period.

Mechanical and Process Problems—

Chemical mixing—D/C stage—As stated in Appendix A, the importance of good mixing of chemicals and pulp in bleaching cannot be overemphasized. This has been properly recognized in the pulp and paper industry for years. Relatively recent published papers(45, 46, 47) germaine to the Chesapeake mixing systems refer to earlier published articles. Chesapeake installed a conventional mechanical dynamic mixer initially and replaced with a static, in-line system for the reasons explained in the next two sections.

Fiscalin mixer—The mixer installed initially ahead of the D/C stage towers was the Fiscalin mixer. This mixer or a similar one was widely used in conventional chlorination bleach plants throughout the industry.

At first chlorine dioxide solution was added to the unbleached pulp ahead of the fiscaline with chlorine added after the Fiscalin and before the short-time retention towers. Not long after startup, the chlorine and chlorine dioxide solutions were combined before adding them to the unbleached pulp ahead of the Fiscalin.

From startup time (December 1972; February 1973) until the mixer was disconnected in March 1976, Chesapeake had no reason to question the mixing efficiency of the Fiscalin mixer. Reviewing the recent literature, Elliott, et al(47) confirmed the Fiscalin to be superior to a wobble plate mixer in a mill scale evaluation. Chen, et al(46) claimed better mixing performance, lower maintenance cost, and lower energy use for an in-line static mixer compared with a dynamic mixer. Swenson(45) discussing the mixing mechanism for in-line static mixers, reported that more than 56 pulp mills had installed in-line mixers to replace mechanical mixers and Swenson described lower chemical use following a "microturbulent" mixed chlorination stage. Chesapeake decided to explore the alleged advantages of the in-line mixer. The main factors studied involved economics of the two systems principally, (a) reduced operating horsepower and (b) no maintenance costs. No organized before and after studies of mixing performance or change in chemical usage were made.

In-line static mixer—An in-line static mixer system was selected based on vendors recommendations. It was installed in parallel with the Fiscalin mixer to use the latter as a spare if necessary. The in-line mixers were installed in January 1976, the Fiscalin was shutdown at the same time and was disconnected but left in place in March 1976.

Installation procedures were recommended in two internal memoranda(48, 49). One(49) detailed the system design and operating pressure losses expected. The same memorandum stated that if the in-line mixer were installed, Chesapeake should realize an annual saving of at least $8000. And with improved mixing, chemical consumption would be reduced. The latter projection has not been documented.

The in-line mixers installed consisted of a 102 mm (4-inch) diameter by 1,007 mm (42-in.) long unit with 6 elements and two 305 mm (12-in.) diameter by 1,830 mm (6-ft) long units (in series) with 6 elements each. The smaller unit mixed chlorine and chlorine dioxide solutions and gases before the combined stream entered the feed end of the larger unit to be mixed with unbleached feed pulp.

Total pressure drop across the large and small assemblies was calculated to be 51 kPa (7.4 psi) and 55 kPa (8 psi), respectively. The net effect, because of other changes, on system pressure drops would be(49) an increase of 26.9 kPa (3.9 psi) across the chemical feed system and a reduction of 26.9 kPa (3.9 psi) on the brown stock system.

Since the in-line mixers were installed in January 1976 maintenance costs have been zero. Only power required has been to pump the combined chemicals into the unbleached pulp; this is small. Therefore, it is certain that savings projected for the in-line mixers based on maintenance costs and energy reduction have been achieved. As mentioned above, no organized study of chemical cost savings has been done but a cursory check of mill operating reports shows no clear change in chemical costs.

Twin roll press—Within less than 10 days of sporadic operation after the initial bleach plant startup on December 3, 1972 the plant shutdown for 2 months because of a major mechanical failure in the twin roll press (TRP). This equipment was the cause of so much down time and process problems that substantial space is devoted in this report to a description of the problems and the solutions developed.

Original specifications called for a drilled 317 ELC stainless cover for the two rolls of the TRP. Instead, these rolls were installed initially with a 317 ELC stainless, woven-wire cover. At press loadings less than the 444 kN/m of lineal width [2500 lb/linear in. (pli)] for which the TRP was designed the woven-wire covers failed by wire fracture.

There being no other way to dewater pulp after the D/C stage to 30 to 35 percent consistency required just ahead of the O-stage, the entire plant shut down for two months until new, drilled plate covers could be installed.

As an aside, both the vendor and Chesapeake had expected the TRP roll covers of 317 ELC to last indefinitely although it was recognized that this was the only TRP in the world operating in this environment. Because of such expectations no spare rolls had been acquired prior to bleach plant startup.

The actual cause of wire failure was never clearly established although flexing fatigue appeared to be the main cause. Studies were made to try to establish the cause and to provide guidance in preventing a recurrence. One study(50) arranged by Chesapeake reported that it was determined that crimp marks in the wire, resulting from the wire screen manufacturer's press-lock system, served as initiation sites for stress corrosion cracks. Cyclic loading to high stress in a chloride environment caused crack propagation by a corrosion fatigue mechanism, and on attaining a critical size, caused a brittle failure in that wire strands. A second study(51) arranged by the vendor reported that the wire mesh had cracked severly and disintegrated. Visual and low-power examinations revealed the presence of cracks and pits on the wires. Such pitting of austenitic stainless steel is typical of corrosion attack by chlorides. Metallographic cross-sections showed a crack had propagated in an essentially straight line and had branched only slightly. These features are typical of corrosion fatigue. However, corrosion fatigue usually starts at a pit at the surface and none was present here. Cracks caused by stress corrosion usually have significant amounts of branch cracking. This was not true in the case of these wires; the fractures were very straight. Additional evidence to support the fatigue theory was the appearance of the wire fractures. Subtle fatigue markings (beach marks) appear to be present on the fractures.

Still another study(52) stated that the wires in the screens possessed austenitic microstructures which are normal for 317 stainless steel. The fracture in each wire was a single transgranular crack.

without branching. There were no corrosion products in the cracks or on the surfaces. The surfaces were free of evidence of corrosion. Failure in the screens resulted from flexing or bending of this reltively hard, brittle alloy. The surface notches acted as stress raisers, which provided focal points for development of the fatigue cracks.

These studies support the conclusion that the woven wire cover failure was due to flexing fatigue. However, sufficient evidence was available from the literature(53)(54) to suspect some contribution from corrosion. Steps taken to try to prevent this are described below.

First, to avoid similar flexing fatigue, the originally specified drilled plate covers were selected as replacement for the failed units. For several reasons, this type of cover was not available in the United States. It was necessary to have the covers fabricated and shipped from Finland. Even with a crash schedule, the TRP rolls with new 317 ELC drilled covers (1 mm holes on 2 mm centers, plate thickness 4 mm) were not installed and started up until February 12, 1973.

Second, Chesapeake decided to accept the vendors' recommendation to install an antichlor system to inject SO_2 after the D/C stage and before the TRP to fix the residual ClO_2 and Cl_2. This procedure was practiced at numerous bleach plants with conventional chlorination sequences. Also, an SO_2 injection system was designed into the Chesapeake D/COD sequence between the D-stage and the D-stage washer, presumably to inhibit corrosion in the washer vat. However, only limited literature(55)(54) refers to this technique and the discussions therein appear to be somewhat contradictory. For example, Tappi Monograph No. 27(55) refers to use of sulfur dioxide as an antichlor to reduce pH to improve brightness stability with no mention of its merit as a corrosion inhibitor. At one point Passinen, et al.(53) in their study of corrosion of austenitic steels in ClO_2 solutions state that the SO_2 gas used in a bleaching plant is easily oxidized to sulfuric acid by the action of ClO_2 (gas), resulting in the formation of a very corrosive sulfuric acid-hydrochloric acid dust. At two other points in their paper they stated that in all (test) solutions, the sulfate concentration was kept constant (0.05 moles/L), when necessary by the addition of Na_2SO_4, since the sulfate ion was found to exert a slight inhibiting effect. Also, if other anions are added to the chloride-containing solution, in general corrosion is diminished. This may be in part because these other anions replace a proportion of the chloride ions adsorbed on the metal surface and thus act as inhibitors. It has been found that the inhibitor effect of the following anions increases as follows:

$$SO_4 < ClO_3 < NO_3 \tag{5}$$

For achievement of the same inhibitor effect, there is needed more of the anion to the left than that to the right.

Whatever the uncertainty of success, Chesapeake installed(56, 57) the injection system to provide an amount of SO_2 ahead of the TRP sufficient to be oxidized to sulfate by a residual of 0.1 to 0.2 g/L of available chlorine after the D/C stage whether from ClO_2 or Cl_2, but most likely the latter. Two months later the record shows that the injection system had not been and was not then working satisfactorily. One memorandum(56) stated that 37 to 74 kg/hour (81 to 162 lb/hour) of SO_2 would be required for an available chlorine residual of 0.1 to 0.2 g/L after the D/C stage. A second memorandum(58) stated that pitting type corrosion on the breaker conveyor (immediately following the TRP) makes the operation of the D/C SO_2 system imperative. To date we have not been able to make the system function properly.

The main problems involved (a) shrinkage, closure, and breakage of PVC pipe carrying the SO_2, (b) plugging of the water line to the ejector, and (c) malfunction of the stream heater to vaporize the liquid SO_2. When these problems occurred in 1973 they were corrected quickly but they did interrupt steady addition of SO_2 ahead of the TRP. Eventually, the system became more reliable but its capacity was insufficient to handle the peak swings in residual chlorine concentration from the D/C stage.

Daily operating logs available for the years 1973—75, inclusive, do not list the SO_2 addition rate by the pre-TRP system. They do show that the SO_2 injection system between the D-stage and D-stage washer added 9 to 45 kg/hr (20 to 100 lb/hr) or 216 kg/day (480 lb/day). Purchasing records(56) show delivery of 85,185 kg (189,300 lb) between December 1, 1972 and February 20, 1974. However SO_2 was injected after the D-stage through February 15, 1975. Production and run time records show 139,750 metric a.d. tons (154,080 a.d. tons) of bleached pulp produced in 614 days or an average production rate of 227 metric a.d. tons/day (251 a.d. tons/day) during the period December 3, 1972 through February 15, 1975. These numbers compute to an average SO_2 use of about 5.8 kg/hr, which seems low but could also be misleading. For example, during many operating shifts, SO_2 addition was zero, especially when the Cl_2 or ClO_2 residuals were quite low to zero.

With SO_2 injection system problems after the D/C stage and widely variable (9 to 45 kg/hr) addition rates after the D-stage, it cannot be concluded from the measurements made and records available whether or not the amount of SO_2 added was adequate, inadequate, or perhaps excessive at times.

Regardless of what the use rate of SO_2 may have been ahead of the TRP, the drilled plate cover failed again between December 31, 1973 and January 8, 1974. Not only the covers failed but the shell supporting the covers failed also. The points of failure were juxtaposed at one end of the roll, the plate cover cracks lying directly above the shell cracks. The rolls were shipped to IMPCO for repair and replacement of both shells and covers. They were returned late in January and the bleach plant started up January 28, 1974.

While in the IMPCO shops, gussets were welded to the ribs of each roll, to strengthen the ribs, although no cracks or bowing were found in the ribs. Also, slingers were installed on each roll shaft to keep filtrate from the bearings.

A letter from IMPCO(59) discussed the causes of failure, suggested acquiring two spare rolls, explored the possibility of Hastelloy C roll covers, re-emphasized the need for a D/C washer and enclosed several documents related to the subject of roll failure. The IMPCO letter also suggested a means to equalize nip loads across the entire roll length, since non-uniform loading was strongly suspected as contributory to failure. Chesapeake installed this system.

Chesapeake acquired spare rolls with 317 ELC covers rather than investing in Hastelloy C rolls, which were estimated to cost $368,400 (1974 dollars) with a one-year delivery time. Chesapeake also contracted with IMPCO on April 25, 1974 to install a titanium, fiberglass vacuum washer between the D/C stage and the TRP to reduce chloride concentration to the press and to reduce cost of NaOH for pH control before the press. The need for such a washer had been anticipated in the original plant layout. The washer was installed in the space provided earlier and started up in July 1975.

The documents enclosed in the IMPCO letter(59) included two(60)(61) which offered the most pertinent information on the failure causes. The first(60) reported that the stainless steel shell, drill plate, and liner of each roll exhibited typical evidence of failure by stress corrosion. The complete absence of cracks on the opposite end of the mating rolls indicates asymmetrical loading to be a contributing factor. The molybdenum content of each component was found to be less than specified, the consequences of which are not predictable for the system's particular stress level and corrosive potential. The supplement(61) to Johnson's first report stated that new evidence concerning chemical analyses described in the report(60) had been brought out. Chemical analyses were done by four methods for molybdenum content of the metals, which the first report showed to be below the specified limit of 3 percent, and which Passenin, et al(53)(54) had shown to be critical in Cl_2 and ClO_2 corrosion of austenitic stainless steels. The first analysis, an atomic absorption method, indicated low molybdenum content. To be more certain, tests were duplicated by another laboratory that used emission spectography. Some differences between the two were so great as to cast doubt on the validity of either. A third laboratory using x-ray flourescence analyses indicated the alloys to be the required Type 317L.

In view of the differences, a wet chemical referee type test (ASTM E-30-70, Method 97 through 101, Book 32, April 1972) was carried out by the laboratory that had used the atomic absorption technique. The results showed molybdenum contents of 3.28, 3.16, 3.05, and 3.14. Thereupon, metal in the roll parts were declared to above the specification limit of 3 percent molybdenum.

At a molybdenum content of 3 percent and above (Figure 6 of the paper by Passinen, et al(53)) shows very low corrosion rates of austenitic stainless steels at 50° C (122° F) and a pH of 2.0. The conditions approximate the actual operating conditions at the TRP. Therefore, it can be reasoned that exceptionally high stress at one end with the asymmetrical loading apparent, was the chief cause of roll failure. Although pits were reported by Johnson(60) at the unstressed end of the roll, there were no cracks observed. This conclusion is supported by experience with the TRP since January 1974.

The rolls, even with SO_2 injection ahead of the TRP described earlier to react with residual Cl_2 and to inhibit corrosion, showed cracks during the week of August 27, 1974. One spare roll had been delivered April 9, 1974; another October 22, 1974. While waiting for the second roll to arrive, the cracks observed in August 1974 were welded. Inspection on October 22, 1974 revealed weld wear but the rolls were still runnable—so the set of spare rolls were not installed. Again in late December 1974 roll inspection showed a worsening of the cracks which were repairable. However, the rolls had to be replaced in July 1975, about the same time as the D/C washer startup with caustic addition for pH control at 4.0 to 4-5. Another set of new spare rolls were delivered between August and December 1975 presently on hand. An inspection August 7 of the operating rolls showed cracking had started again in the drilled plate cover. However, to date neither covers or rolls have been replaced.

For these experiences it appears that stress rather than corrosion is strongly contributive to roll cracking. There may be some question as to how well the rolls were protected with SO_2 injection from February 1973 to January 1974; from then until August 1974 and thereafter to July 1975. However, after the last date mentioned the rolls should have been reasonably protected from a corrosive atmosphere with the D/C washer and pH control. Cracking observed recently then is probably due largely to stress.

Recurrently the TRP lost production capacity because the drilled holes in the 317 ELC cover partially plugged. Initially, a metal doctor blade operating against the external surface cold-worked the metal part-way across the hole opening. This was clearly shown by sectional photomicrographs taken by Johnson(60) in February 1974 when examining the roll cover for cause of failure. This problem was solved by shifting to FRP doctor blades in the TRP.

At various times the TRP vat and roll cover, including the drilled holes scaled significantly. The scaling material first observed in February, March and April 1974 was barium sulfate which built in the vat to a thickness of 6.3 mm (¼-in.). When this became heavy enough to block the drilled holes so as to reduce production, the only really successful scale removal methods used were chipping on the vat surfaces and high pressure water jetting of the drilled holes. After the first experience with $BaSO_4$ scale several additives were tried ahead of the TRP to prevent the scale deposit forming. As so often happens in complex pulping and bleaching systems scale deposits come and go without known or observed reasons. Therefore, no firm conclusion was ever reached regarding efficiency of a particular additive at that time.

During the three-month period (February, March, April, 1975) before installation of the D/C washer in July 1975, barium sulfate scale appeared again. The TRP and drilled holes were cleaned as before.

After the D/C washer installation, when pH to the TRP was controlled to 4.0 to 4.5 with caustic addition, scaling was noted again on October 14, 1975 on the D/C washer, the TRP and the O_2 washer repulper. Scale was sampled from the riser trumpets on November 25, 1975 and from the ribs of the TRP on December 9, 1975. This scale was determined to be calcium oxalate. The TRP holes were plugging on December 17, 1975 so the TRP was acid washed with HCl and water jetted to clear the drilled holes.

Calcium oxalate scale appeared again on the D/C washer and TRP March 5, 1976. Chesapeake, searching for the cause, recalled three facts which could explain recurrence of the oxalate scale. First, it was observed during the RO laboratory and field trials (EPA Grant No. R803525-01) by The Institute of Paper Chemistry on Chesapeake's bleach plant effluent, that the O-stage effluent contained relatively high calcium and total oxalate concentrations.(62) Second, before installation of the D/C washer, pH to the TRP had been as low as 1.5 to 2.0, whereas after the D/C washer was operating pH was controlled to 4.0 to 4.5 with caustic. Third, oxygen stage filtrate was recycled to the repulper on the D/C rewasher to reduce caustic consumption. Under conditions of higher pH and increased oxalate concentrations at the washer it was not surprising that calcium oxalate scale formed in parts of the D/C washer and in the TRP. The O-stage filtrate recycle to the D/C washer repulper was stopped. Since then calcium oxalate scale has not recurred.

Very soon after the bleach plant started up again in February 1973 with drilled plate covered rolls in the TRP, three additional problems were encountered. For orientation Figure C—10 shows sectional views of the TRP.

The risers to the vat of the TRP plugged at positions A. The vat itself plugged at area and positions B. Both problems reduced production rate or caused downtime. The third problem was excessive fiber loss in the pressate. The last two problems appeared related since vat plugging caused higher velocities through the drilled holes, slowed roll rotational speed and increased fiber loss. The fiber loss problem will be discussed first.

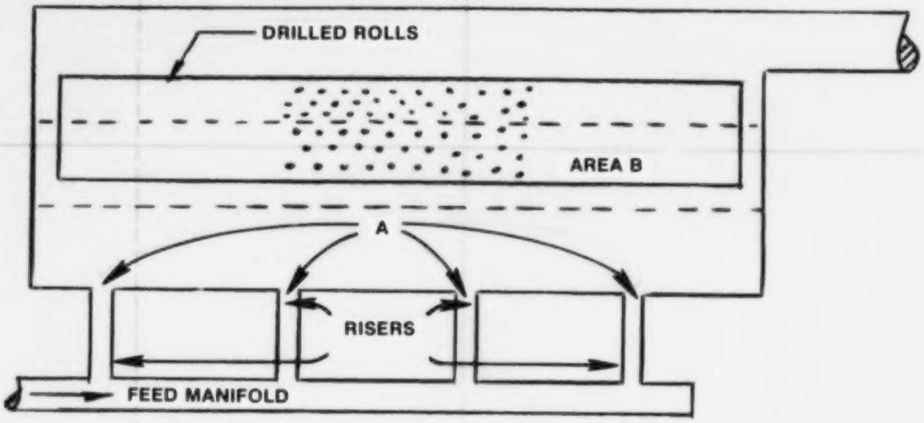
Fiber loss from the TRP was first mentioned in the Supervisors' Meeting on March 27, 1973. At a 272 metric tons/day (300 tons/day) production rate, fiber losses to the sewer approached 9 metric tons/day (10 tons/day). Technical and Engineering Departments were directed to cut this fiber loss at the earliest possible time. Subsequent tests showed the fiber losses to be much higher under certain conditions.

Although three screen (saveall) vendors offered equipment for trials to recover TRP fiber loss, long delivery time was projected for both saveall and the pulp required. Therefore, during March and April 1973 an intensive study of TRP operating parameters was undertaken to minimize fiber losses.

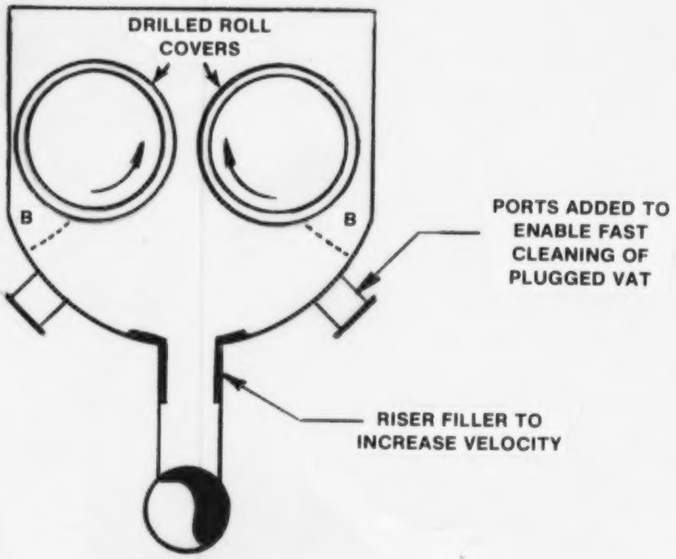
Table C—10 lists the fiber length spectrum in the feed stock to the TRP compared with that of fibers in the TRP pressate.

These data were taken when the fiber loss from the TRP was 2244 mg/L (18.7 lg/1000 gal) or 16 tons/day, and production rate was 232 metric tons/day (256 tons/day) at 3.5 percent consistency of the feed stock. They show the TRP to be non-selective in fiber retention except that the pressate appeared to have a somewhat higher fraction of the shorter fibers which passed a 100-mesh screen. This was not a surprising result.

To establish a reasonable target fiber loss Chesapeake's TRP fiber losses were compared with fiber losses from systems of other mills.(64) Table C—11 shows suspended solids (SS) losses from four other mills.



SIDE VIEW


DRILLED ROLL COVERS
B
B
PORTS ADDED TO ENABLE FAST CLEANING OF PLUGGED VAT
RISER FILLER TO INCREASE VELOCITY

END VIEW

Figure C—10. Side- and end-view schematics of the twin roll press (TRP).

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The fiber losses shown in Table C—11 range from 3.4 to 8.5 kg/metric ton (6.9 to 17.7 lb/ton) at 40 to 75 percent bleached hardwood in the four mills. These losses convert to those shown in Table C—12 at different daily pulp production rates.

TABLE C—10. FIBER LENGTH OF FIBERS LOST THROUGH THE TWIN ROLL PRESS(63)

Screen mesh	Equivalent fiber length (mm)	Fiber content Feed stock (%)	Fiber content pressate (%)
14	3.1	1.3	1.1
20	2.5	14.9	10.9
48	1.2	41.9	40.1
100	0.69	20.3	20.5
100	< 0.69	21.6	27.4

TABLE C—11. SUSPENDED SOLIDS LOSS AT FOUR MILLS

Mill No.	Suspended solids Kg/metric ton	Suspended solids lb/ton	Percent hardwood bleached
A	8.9	17.7	75
B	4.4	8.7	65
C	3.4	6.8	60
D	3.9	7.7	40

TABLE C—12. FIBER LOSSES (TABLE C—11) CONVERTED TO METRIC TONS/DAY

Production rate Metric tons/day	Production rate (tons/day)	Production rate kg/metric ton	Fiber losses Metric tons/day	Fiber losses (tons/day)
181	(200)	3.4—8.9	0.62—1.61	(0.68—1.78)
227	(250)	3.4—8.9	0.77—2.02	(0.85—2.23)
272	(300)	3.4—8.9	0.93—2.42	(1.03—2.67)

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These losses were compared with the range of losses indicated in Figure C—11. The range of observed concentrations in the TRP pressate shown on Figure C—11 from (a) the average of about 400 mg/L (3.3 lb/1000 gal) for a production rate of 181 metric tons/day (200 tons/day) to (b) almost 3500 mg/L (29 lb/1000 gal) at 236 metric tons/day (260 tons/day), corresponds, at 3.5 percent feed, consistency respectively, to 1.9 and 23.4 metric tons/day (2.1 and 25.8 tons/day).

Clearly, both are far above the losses reported for the four mills in Table C—12. Furthermore, these mills measured SS by filtering pressate through filter paper. This method removed essentially 100 percent of the SS and would tend to give higher loss values (SS_F) than the method normally used by Chesapeake which measured the fiber (SS_w) retained on a 150 mesh wire. Figure C—12 shows the correlation between the two methods. At SS_w = 400, SS_F is 650 mg/L or 62.5 percent. At SS_w = 1305, SS_F is 1590 mg/L or 21.8 percent higher. This indicates Chesapeake's method was very conservative on the low side and that Chesapeake's fiber losses from the TRP was far above that from the other mills. Therefore, the decision to study TRP process parameters to reduce fiber loss, pending installation of a fiber recovery system (discussed below), was well justified.

Since the TRP is basically a fiber dewatering device it would be expected that pressure drop across the vat and related pressate velocity through the drilled holes would strongly influence fiber loss from the TRP. Figures C—11, C—13, C—14, C—15, and Table C—13 taken from a report(65) on the TRP study show the relationships among TRP vat pressure, drive speed, production rate and fiber loss.

Figure C—11 shows that fiber loss increased as vat pressure increased. The fiber loss increased at a constant vat pressure when production increased from 181 to 236 metric tons/day (200 to 260 tons/day) (Lines I and II). The loss increase was due probably to some vat plugging, rather than production increase entirely, since the Line II data were taken 24 hours after the Line I data. Plugging had been observed earlier to occur within one hour after cleaning; certainly some plugging had occurred in 24 hours. Line III shows that fiber loss increased sharply when the vat was plugged more severely. After the data shown by Line III were taken the TRP was shut down to check extent of plugging. The area available for drainage had been reduced from 215° of te circumference to 70 to 80°. Also, a single data point at a production rate of 2650 tons/day indicated even more severe vat plugging.

Generally, it was observed that most of the very high fiber losses above concentrations of 1200 mg/L (10 lb/1000 gal) or almost 43 kg/metric ton (86 lb/ton) of pulp, occurred when the vat was plugged to some degree. This situation was corrected eventually as described shortly. In the meantime, however, it was necessary to find some mode of operation that would reduce fiber loss as much as possible.

To achieve 120 mg/L (1 lb/1000 gal) (4 kg/metric ton), Figure C—11 Line I shows it would be necessary to operate at a vat pressure of 41.3 to 48.2 kPa ga (6 to 7 psig). At this vat pressure the production rate was only 181 metric tons/day (200 tons/day), far below the design of 272 metric a.d. tons/day (300 a.d. tons/day) average.

Seeking a better understanding between fiber loss from the TRP and production rate, the data shown in Table C—13 were collected(65) at random. The relationship between loss and rate is not clearly apparent from Table C—13 but emerged roughly to show higher fiber loss at higher production rate when the data were plotted as Figure C—14. Obviously, the data scatter was wide and was attributed to unknown degree of TRP plugging although it does appear that the TRP was relatively clean when the fiber contents in the pressate are compared with the much higher losses shown on Figure C—11.

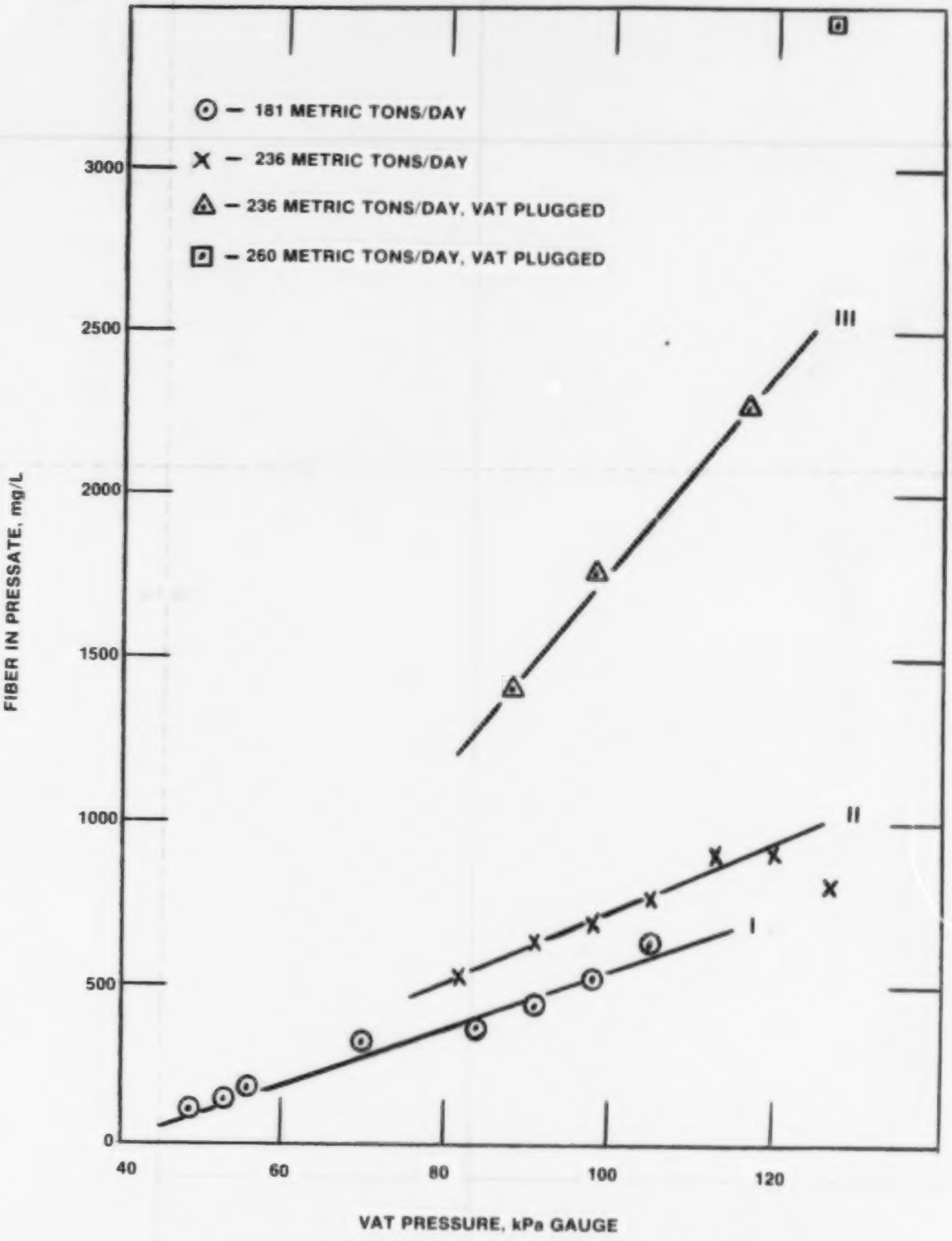

Figure C—11. TRP fiber losses as function of vat pressure.

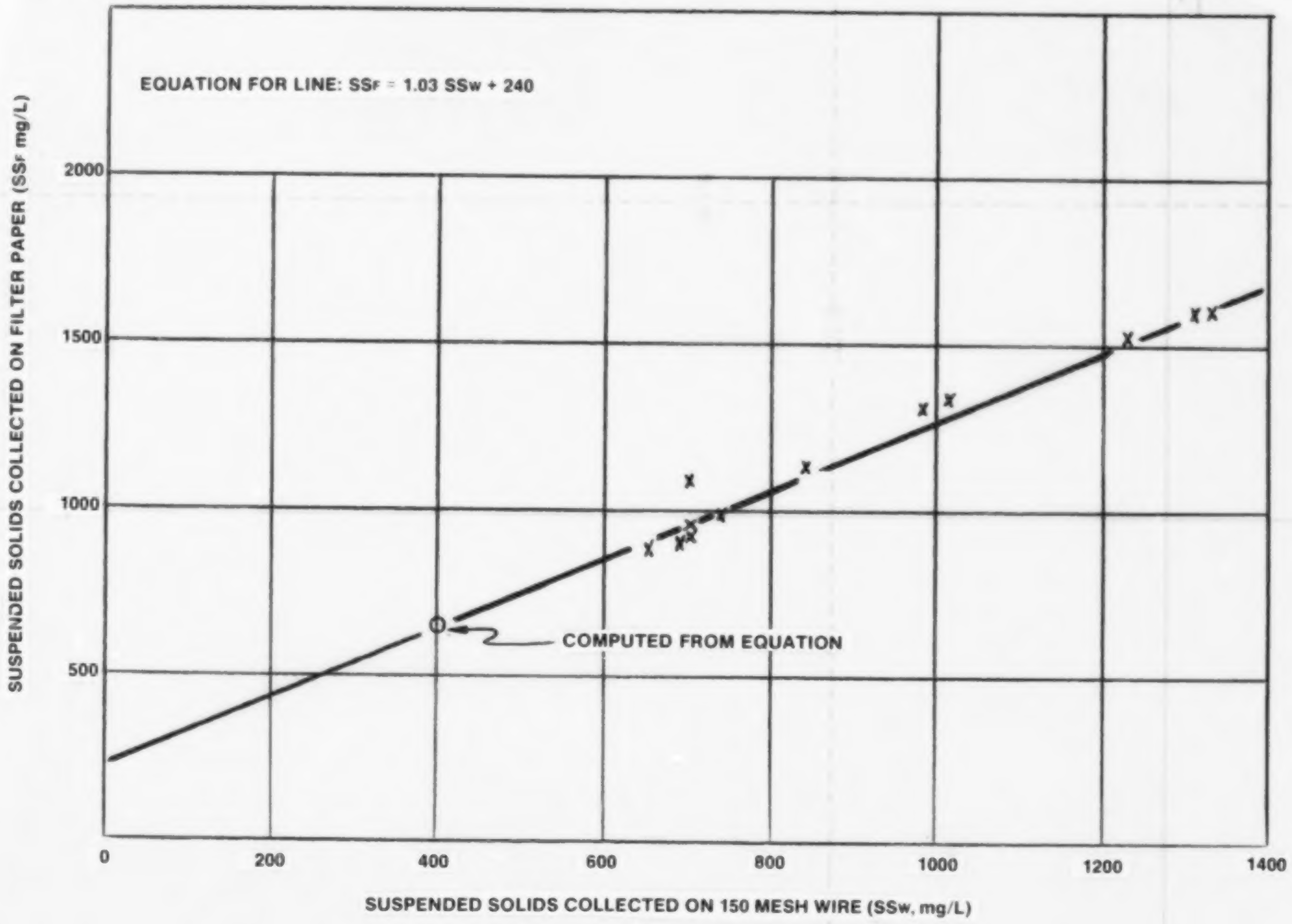

Figure C—12. TRP pressate suspended solids measured by filter paper method vs. solids collected on 150 mesh wire.

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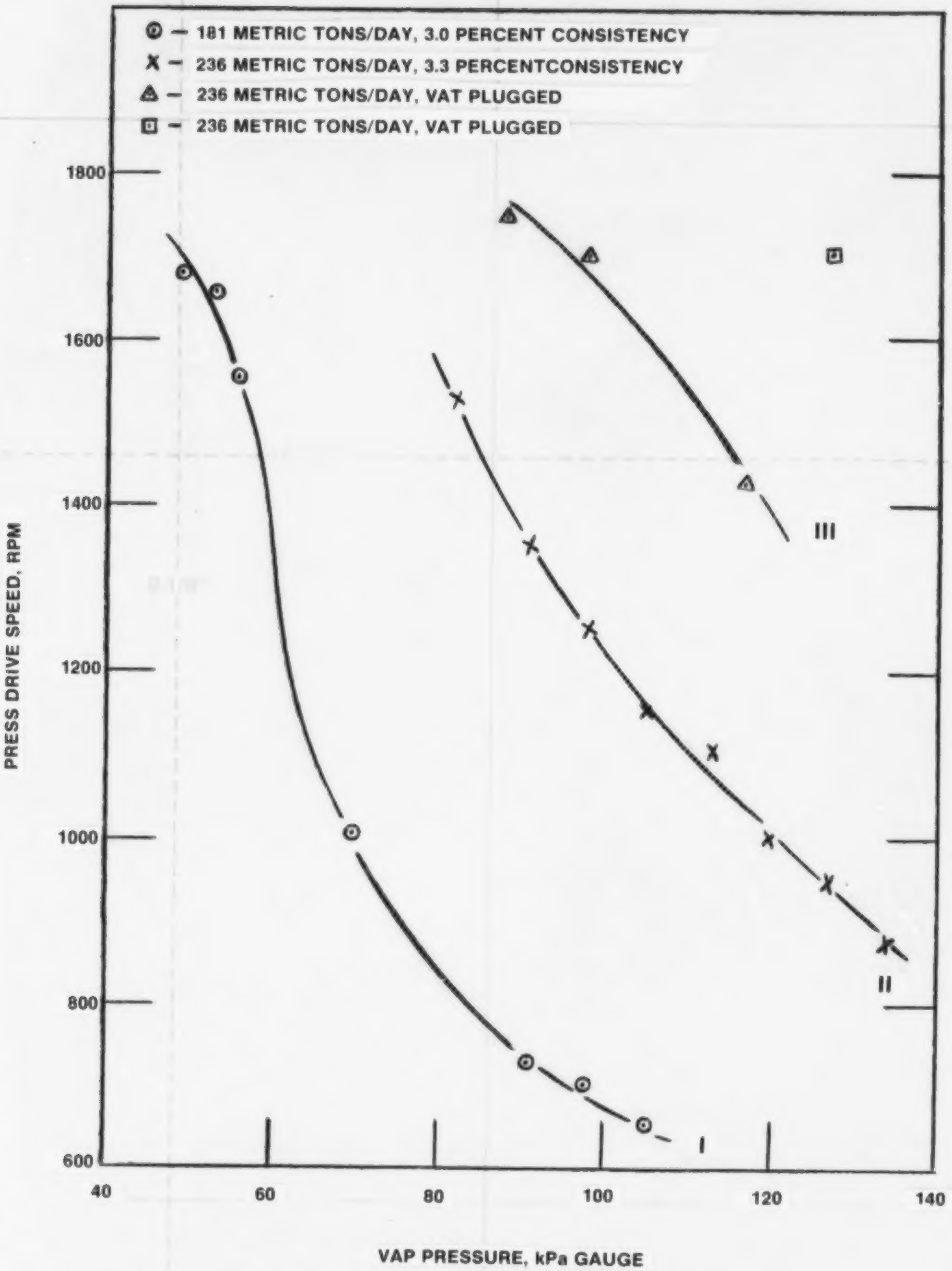

Figure C—13. Drive speed vs. vat pressure for TRP.

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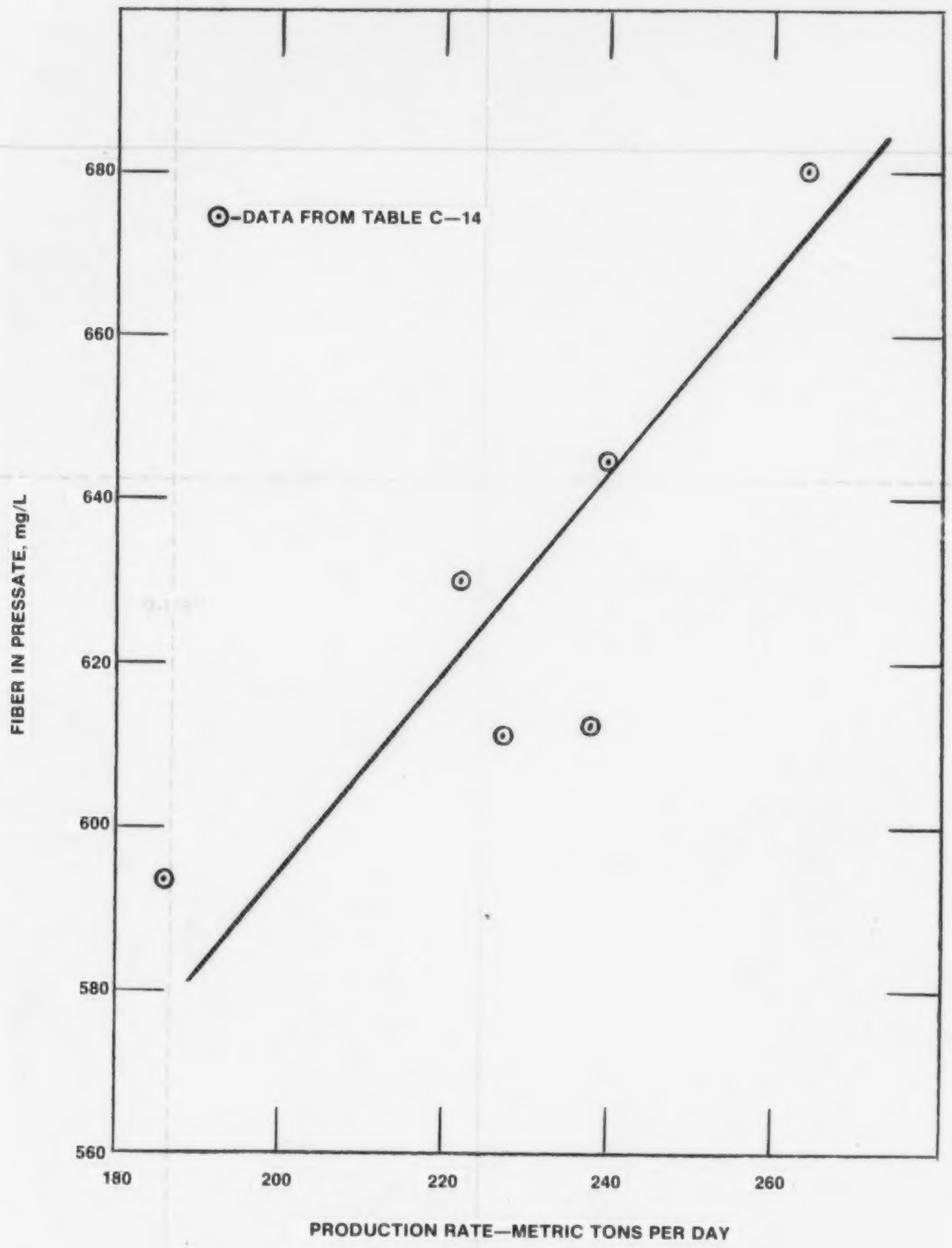

Figure C—14. Relationship between fiber content in pressate and production rate for the twin roll press.

A more systematic study of pressate fiber content and production rate was made at a constant vat pressure of 63.7 kPa (9.25 psig) with a vat known to be clean at the time the data were taken. Feed consistency and flow rate were measured carefully and the effect of each established by multiple regression analysis. The original data from this study could not be located but the following equation and correlation coefficient were reported(35).

$$\text{Drive Speed (rpm)} = 0.00274\,(Q_E)^{1.67}\,(C)^{0.94} \tag{6}$$

$$\text{Drive Speed (rpm)} = 30.365\,(Q_M)^{1.67}\,(C)^{0.94} \tag{7}$$

where Q_E = gpm;
C = percent consistency
and Q_M = m³/min

$$\text{Correlation coefficient} = 0.965 \quad (8) \tag{8}$$

Assuming constant 3.5 percent consistency and converting Q_E, Q_M, and C to production rate, Equations (6) and (7) become

$$\text{Drive Speed (rpm)} = 0.123\,(PR_E)^{1.67} \tag{9}$$

$$\text{Drive Speed (rpm)} = 0.1416\,(PR_M)^{1.67} \tag{10}$$

where PR_E = short tons/day
and PR_M = metric tons/day

TABLE C—13. EFFECT OF PRODUCTION RATE ON FIBER LOSS

Production rate		Fiber loss	
Metric tons/day	**tons/day**	**mg/L**	**lb/1000 gal**
186	205	593	4.94
222	245	630	5.25
227	250	611	5.09
238	262	612	5.10
240	265	644	5.37
264	291	680	5.67

Vat pressure = 63.7 kPa (9.25 psig).

Figure C—15, Line I is a plot of Equation (10) at a 3.5 percent consistency. It shows clearly that at the top drive speed of 1750 rpm the production rate through a clean TRP can be no more than 275 metric tons/day (304 tons/day) on an oven dry basis.

This corresponds to 306 metric a.d. tons/day (338 a.d. tons/day) which is close to the design peak rate of 272 metric a.d. tons/day (300 a.d. tons/day). Based on probable downtime for maintenance, repair and production priorities it could be projected that the peak rate would have to be substantially higher than 306 metric a.d. tons/day to achieve the design average of 272 metric a.d. tons/day (300 a.d. tons/day) in a normal 350 day year. The only apparent way to boost the peak production rate was to operate the TRP at higher than 3.5 percent consistency, at higher than 1750 rpm drive speed or both.

It appeared then that Chesapeake would be production limited by the TRP, unless the undesirable fiber loss could be solved independently, vat plugging could be reduced in some manner at feed consistencies higher than 3.5 percent and drive speeds could be increased.

This left Chesapeake in the position of (a) declaring the TRP undersized if operation in a partially plugged mode was normal, (b) installing as soon as possible a system to recover fiber from the TRP pressate and/or (c) finding a way to stop or reduce the plugging to increase production rate and reduce fiber loss.

Striving first to correct the problem of fiber loss from the TRP, Chesapeake studied the three different savealls mentioned earlier. The one selected was a single disc type which dramatically reduced TRP fiber loss. However, this system was by no means trouble-free and did not correct the system fiber loss problem entirely. For example, there were occasional overflows at the saveall that went to the sewer. To correct this, the system was piped and re-piped so that the fiber recovered by the saveall, as well as the infrequent overflows at the saveall, returned to the brown stock chest ahead of the D/C stage. With this system, fiber loss was still about 3 tons/day in the filtrate from the saveall and the fiber recovered from the TRP was re-bleached one or more times. The latter was probably not a significant problem and whether or not it was became academic in the Chesapeake system when the D/C washer was installed in July 1975. The saveall was shutdown and all pressate from the TRP was recycled to the vat of the washer where all fiber was recovered.

Although the TRP fiber loss problem was solved at Chesapeake as just described, any other TRP installed should be checked carefully for excessive loss of fiber. Also, the TRP may have another operating characteristic that could dictate the method adopted to recover pressate fiber loss. For example, recycling the pressate to the D/C washer vat has consistently caused poorer D/C washer performance. First, the mat formed on the washer non-uniformly. Second, the displacement ratio dropped sharply and the consistency of the pulp mat leaving the washer dropped also (11). These symptoms are typical of washer performance when air is dispersed in the liquid phase and/or occluded on the fibers in the water-fiber mixture in the washer vat. Air was, and still is, observed in the pressate. Thus, even though diverting the pressate to the washer vat saved fiber it was not the ideal solution.

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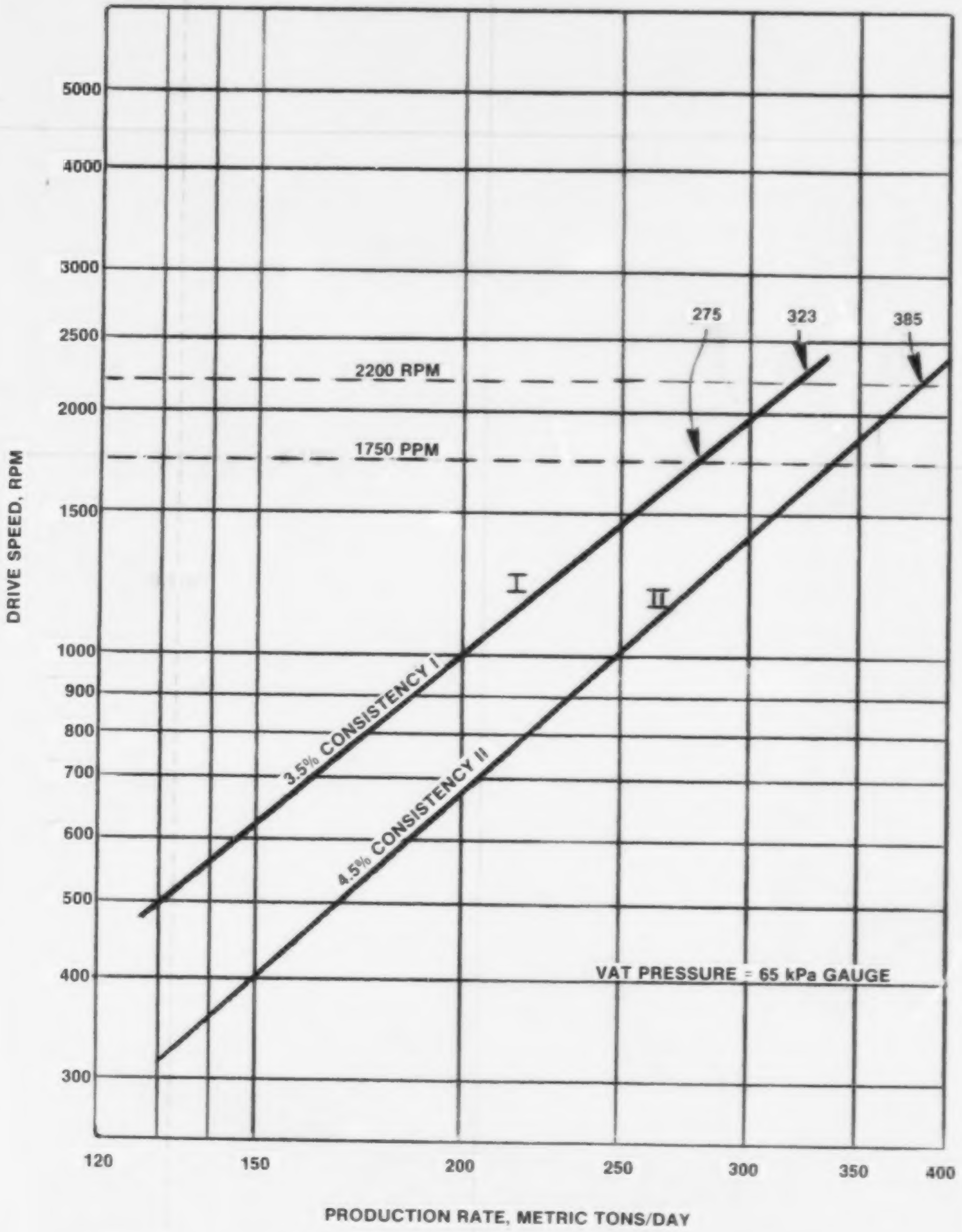

Figure C—15. Effect of drive speed on production rate at constant vat pressure.

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It can be surmised that the TRP performs like the feed end of a low pressure flotation unit. As the water fiber mixture is pumped to 68.9 to 103 kPa (10 to 15 psig) more air shifts from gas to liquid phase. As the liquid phase, or pressate with fine fiber, passes through the drilled holes to atmospheric or vacuum conditions, the air dissolved under pressure evolves as microscopic bubbles at the vast number of nucleation points provided by the fine fibers. In fact, the pressate has been observed to be milky, like an emulsion. There would appear to be no simple way to separate the air and fiber under these conditions.

With respect to TRP vat plugging, as already mentioned, Figures C—11, C—13, C—14, and C—15 show the strong effect it has on vat pressure, fiber loss, production rate and drive speed. While seeking to solve the TRP fiber loss problem some changes were made in the TRP to reduce and, if not eliminate, at least manage, the vat plugging problem.

Plugging of the risers, shown in Figure C—10, positions A, was noted first. Vat plugging was observed thereafter to occur most frequently and seriously above plugged risers. Recalling(65) that another mill had solved vat plugging by dropping TRP feed consistency to 2 percent (and, thus, increased flow volume) Chesapeake concluded it was necessary to increase flow velocity at the riser outlet to the vat. Trumpets designed as riser inserts are shown in Figure C—10, End View, dark lines. These were secured in place by a cross member and eye bolt attachment.(66) These helped reduce the rate of vat plugging but did not solve the problem. However, shutdown frequency and duration for vat cleaning was reduced.

In view of the influence of vat plugging on vat pressure, drive speed, and production rate Chesapeake took other steps. First, the clean out port shown in Figure C—10, end view, was installed to expedite inevitable vat washing. Second, it was decided to operate the TRP at the top drive speed of 1750 to maximize the production rate at any degree of vat plugging. Third, the nip pressure was controlled at a constant level with added protection which would open the rolls nip should a lump of fiber pass between the rolls. This last mentioned step was taken to prevent extreme local stress on the drilled rolls, a condition which might cause roll cover cracking to start. Fourth, when the system(59) to ensure equal nip loading was installed, an audible alarm was included to alert the operator to significant unequal roll loading which might develop. Fifth, after drive motor failure in January 1975 because of overload and overspeed, the motor was replaced in kind but a search was started for another motor that would operate at higher speed (see Figure 15) and higher power input. Such a motor was located and the 125-hp, 1750-rpm motor was replaced on January 21, 1976 with a 150-hp, 2200-rpm motor.

Operating at higher consistency would tend to accelerate vat and riser plugging because of lower vat velocities. In fact, as already mentioned, one company reported(65) that the only solution it found to vat plugging was to reduce consistency to almost 2 percent. This meant higher flow at a given production rate and, in turn, provided higher velocities and greater turbulence in the vat. Chesapeake speculated that the riser trumpets would increase vat velocity enough to enable operation at consistencies higher than 3.5 percent. Figure C—15, Line II shows a projection for production rate at 4.5 percent consistency. The equation for this line is:

$$\text{Drive Speed} = 0.1118\ (\text{PRM})^{1.67} \qquad (11)$$

At a drive speed of 2200 rpm the production rate was projected to be 323 metric o.d. tons/day (356 o.d. tons/day at 3.5 percent consistency and 383 metric o.d. tons/day (425 o.d. tons/day) at 4.5 percent consistency. When converted to air dry tons both projected rates were well above the design peak rate of 299 metric a.d. tons/day.

Summarizing this section on TRP problems, this equipment, in the first commercial application of this type, caused extensive shutdowns of the bleach plant because of press roll cover failure and both plugging and scaling in the risers, vats, and drilled holes. Fiber loss was also excessive. In time, these problems were solved or changes made in the TRP system so that operating conditions were tolerable.

When Chesapeake decided, after the grant period, to expand production in the bleach plant and elsewhere in the mill, a second TRP was installed. Recognizing from earlier experience that press roll and vat cleaning would be necessary, the second TRP was installed at an elevation higher off the floor to make such cleaning easier. The movable press roll moves in an arc toward the fixed roll to maintain better parallelism between rolls. Furthermore, the movable roll is loaded at each end by an air bellows instead of the horizontal, threaded adjustors as in the first TRP. This should improve responsiveness to equalize unequal nip loading. The longitudinal seals were made easier to clean and replace. Finally, the overall mechanical structure was built more ruggedly to withstand the high and prolonged stresses applied continuously to the press rolls.

None of the changes in design could be construed as steps to reduce the fiber loss characteristics of a TRP, nor are there any steps other than post-TRP fiber recovery that Chesapeake can recommend.

Oxygen reactor—The oxygen reactor, consisting of a steel outer shell lined inside with Carpenter 20, has given exceptional, trouble-free service. Corrosion, feared at first, has not appeared. Also, no problems with the lining have occurred.

However, several problems with the fluffer, the discharge device, and the drive for the latter have caused considerable downtime. Chesapeake or the vendor corrected these problems, or replaced the equipment one or more times so that eventually these components of the oxygen reactor system required no more than normal maintenance. Furthermore, operating parameters of the reactor have been changed numerous times. The problems, their correction, and the comparison of operating parameters at startup and at the present time are detailed in the following paragraphs.

There have been no serious problems with the safety systems on the oxygen reactor, however, more discussion is included here because of the importance of its design features and functions. Appendix A describes the mechanical nature of the systems. Reviewing the reference section will enhance understanding of the following discussion.

Various parts of the safety systems are designed to reduce excess pressure or temperature in the reactor or to quench any fires. Fires may start because of (a) very high pulp consistency (38 to 40 percent) and the exothermic reaction of O_2 and pulp in the bleaching reaction or (b) small fragments of metal enter the reactor with the pulp. The water spray system operates automatically to quench any fires.

The principal fires in the reactor to date occurred near startup in 1972 and 1973 when (a) the lineal nip pressure of the TRP almost reached the maximum of 444 kN/m (2500 pounds per lineal inch) and (b) when the woven cover on the press rolls of the TRP disintegrated allowing 317 stainless steel fragments to reach the reactor. One other fire occurred since startup when a fine mesh screen at the O_2 inlet to the reactor ignited. The water spray system quenched all three fires automatically.

Only one rupture disc has blown in the six years since startup. This was caused by an accidental over-pressurization with 1,034 kPa (150 psig) steam fed directly to the reactor.

The remotely operated vent valve has been used numerous times to reduce pressure and or temperature caused by miscellaneous malfunctions in the various equipment feeding the reactor contents. A typical instance is when more than the normal caustic at 7 percent (or higher) concentration is added at the steam mixer. In this case the reaction proceeds more rapidly in the reactor and the temperature rise is greater than usual.

The vent valve is used often also to reduce inerts in the gas phase over the pulp bed in the reactor. Dropping the operating pressure and changing the O_2 injection points appears to have reduced the venting frequency required to maintain the desired O_2 purity; the reason is not clear.

Correct functioning of the fluffer is important for uniform bleaching in the oxygen reactor. Soon after initial startup in late 1972 and early 1973, the bearings and seal of the fluffer had to be replaced. Because of mechanical design of the fluffer, which required removal from the reactor and extensive disassembly, this caused an inordinate length of downtime.

After careful inspection by Chesapeake and the vendor the fluffer was redesigned to make bearings and seals more easily accessible. Since then, the normal wear of these parts has caused infrequent fluffer malfunctions which can be corrected quickly.

Several problems plagued the discharge devices (and the drives rotating them) installed to discharge the pulp bleached in the oxygen reactor. The discharge device installed initially is shown schematically in Figure A—2; the replacement device, still being used, is shown in Figure A—3. The devices and the drives for them are described in Appendix A.

The first problem in the reactor system causing excess downtime of the bleach plant occurred in the hydraulic drive on the initial discharge device. Apparently, due to periodic or steady overload stress, metal particles spalled, or sheared, from parts of the hydraulic motor. These were picked up in the oil flow system and eventually plugged the flow system at one or more points. This shut the system down. Disassembly, flushing the system and re-assembly were time consuming and caused substantial production loss.

After three such instances, Chesapeake decided to replace the hydraulic drive system with a mechanical one. Action started in September 1973, and a variable speed d-c motor with a combination helical and worm gear system was installed late in 1974. Delivery of parts at that time was 8 to 12 months.

The new drive worked well but other problems arose soon after installation. A bronze bull gear on the discharger shaft broke several times due to defects or normal wear until it was replaced with a steel gear in 1976.

The next problem was not the result of a mechanical malfunction per se, but the event caused considerable mechanical damage. An upset in operating conditions in the summer of 1976 caused the bottom of the oxygen reactor to fill with pulp. The discharge device, or table, was lifted from the peripheral trunions on which it normally rested and turned. As the pulp was removed from the bottom of the reactor, the discharge table dropped unevenly and jammed against the reactor housing. Both table and gear box were substantially damaged but repairable during an extended outage of the bleach plant.

In the fall of 1976, the initial discharge table was replaced with the rake type presently used. This new type rake offered the options of operating at longer (almost double) retention times in the reactor or increasing production rate at the original retention time of 30 minutes for 272 a.d. metric tons/day (300 a.d. tons/day). Furthermore, maintenance of the rake type discharger has been negligible.

At startup of the bleach plant, high-purity oxygen entered the reactor just above the discharge table through four 102 mm (4-inch) ports located 90 degrees apart around the reactor. In this position, the ports were at the bottom periphery of the high consistency bed of pulp. No doubt some oxygen passed with the pulp into the dilution zone below the discharge table and, from there, through the 76 mm (3-inch) pipelines which served as pressure equalizers between the top and bottom of the reactor. However, these pipelines frequently plugged with pulp so that it is uncertain how much oxygen reached the gas phase at the top of the reactor via this rouge.

During troubles with the oxygen washer, O_2 gas was detected in the dilute pulp leaving the reactor. This meant some O_2 wastage with the injection points located just above the dilution zone in the reactor. Although never proven, Chesapeake also suspected that O_2 gas channeled to some degree when the gas entered the reactor at the bottom periphery of the pulp bed. For these various reasons, Chesapeake re-located the oxygen injection points to the top of the reactor.

Since the initial startup in December 1972 and again in February 1973, a number of parametric combinations have been expored in the oxygen reactor. There is little merit in researching production logs to extract the incremental changes since the present parameters provide O-stage bleaching equal to those used at startup. Table C—14 shows the set of operating parameters used at startup, and sometime thereafter, and the set being used in 1978.

Oxygen reactor blow tank—The blow tank following the oxygen reactor is described in Appendix A.

Dilute pulp slurry (5 percent consistency) flows continuously to the blow tank, where it is agitated and diluted further before being pumped to the oxygen washer vat. Shortly after the second startup in February 1973, a problem occurred which triggered the need for lining repair and maintenance over a number of years.

An internal memorandum(67) reviews the history of the blow tank problems for early 1973 to May 1977.

TABLE C—14. COMPARISON OF OPERATING PARAMETERS IN THE OXYGEN REACTOR (STARTUP VS PRESENT)

Parameter	Startup	Present
Temperature °C	116	127—132
°F	(240)	(260—270)
Pressure, kPa	620—689	26.9—28.9
(psig)	(90—100)	(55—60)
Consistency, (percent)	28	28
Retention time, min at 272 metric a.d. tons/day (300 a.d. tons/day)	30—45	60*
Caustic added, kg/metric ton	32.3	32.3
(lb/ton)	(65)	(65)
Oxygen added, kg/metric ton	15.9	15.9
(lb/ton)	(32)	(32)

*Optional—depends on portion of reactor bottom used for pulp dilution.

Early in 1973 the dilution pipe broke loose and was whipped around by the agitator, causing numerous cracks and gouges in the lining. The damage was repaired and the tank returned to service.

In December 1974 the bottom lining cracks were again repaired and the steel tank bottom was found to be "springy" and in need of grouting. The grouting was not done.

In September 1976 cracks in the bottom were again repaired. An inspection in December 1976 revealed that the bottom lining was again beginning to crack. Also, minute cracks were evident in the side walls two to four feet above the bottom. It was quite possible these cracks were caused by the "springy" bottom and the fact that previous repairs involved overlaying the existing lining resulting in too great a thickness.

An investigation into various lining systems indicated that flake linings give the best service and require the least maintenance, as well as being the least expensive to install.

Accordingly, in June 1978 the blow tank was relined with flake lining. As this section of this report was written (December 1978) not enough time had elapsed to judge whether or not this tank lining problem has been solved.

Oxygen stage washer—Mechanically, the oxygen stage washer itself has performed well. However, two process problems caused trouble and one, or both, of these may have been partly caused by mechanical malfunction of equipment associated with the washer. These problems and their final solution are discussed in the next two sections.

Without a mat of good uniformity and final proper thickness of about 19 mm (¾-inch) formed on the vacuum washer, poor washing of the pulp occurs. Poor mat forming on the O-stage washer was observed not long after the second bleach plant startup in February 1973. The mat formed was irregular and quite thin, probably no more than 3 to 6 mm (⅛ to ¼-inch). Pulp consistency off the washer was 12 percent or lower.

Numerous factors could cause such performance. Dispersed air (or gas phase) or foam in the fiber-water slurry going to the washer vat and low washer vacuum could be prime reasons. Improper vat level and drum rotational rate could be other causes.

Failure of the valving arrangement inside the drum to maintain a tight seal could cause low vacuum. Likewise, entrained air in the drop leg from any source could cause low vacuum. Failure of the filtrate, or seal, tank to de-gas the filtrate before it recycled to dilute pulp ahead of the washer vat could be such an air, or gas, source.

All these causes are mentioned because they were found to be contributors to the poor O-stage performance after intensive study by Chesapeake and the vendors. Chesapeake discovered two other facts during the study that were fundamental to solving the problem.

First, the filtrate tank appeared to be doing a poor job of de-gasing the filtrate and a superior job of producing enormous amounts of foam continuously. Second, dissolved solids in the filtrate were found to be unusually high, i.e., about 2 percent versus 0.75 percent expected from the system design. The cause was found to be a lower than normal amount of fresh or recycled water entering the washer system, thereby, allowing no significant purge of solids from the system.

In common parlance, the dilution factor was zero or negative rather than positive and the displacement ratio was substantially less than unity. Under these conditions were (a) less new water came into the system with the 28 to 30 percent consistency pulp from the oxygen reactor than the (b) water leaving the system with the 12 to 16 percent consistency pulp, the build-up in dissolved solids had to occur. When this happens in a pulp and paper mill process stream, extreme foaming and tenacious entrainment of small gas bubbles can occur.

The problem was improved dramatically by increasing the shower water on the washer so that the dilution factor was positive and adding fresh water to the seal tank as described below.

Malfunction, or more properly, a mechanical failure in the filtrate, or seal, tank caused some part of the problem described in the previous section. It was well established that the seal tank foamed badly and that foam overflowed to the sewer continuously. Trials were made to reduce the foam with defoamers with moderate temporate success. A spray was designed to apply defoamer continuously at the top of the seal tank but was never installed. Although, it was not clear how and when it happened, stilling and diversion baffles in the seal tank were found to be broken, allowing short-circuiting to occur. This, of course, can shorten retention time and reduce the chance for normal filtrate de-gassing.

The baffles were repaired when found broken but this relieved the basic problem only slightly. Not until steps were taken to reduce dissolved solids and to correct vat level and drum rotational rate as recommended by the vendors did the problem correct itself.

As a part of the basic solution, the seal tank was set up to control at a constant level by addition of fresh water. This helped to purge the system of dissolved solids since overflow from the seal tank was increased. To conserve water, before this overflow went to the process sewer it was used in (a) the fume condenser on the SVP system salt cake filter, (b) the fume scrubber on the D/C washer hood, (c) the main fume scrubber on the oxygen and D-stage washer hoods, the TRP hoods and excess fumes from the D-tower scrubber and (d) in the showers installed in the sewer foam enclosures.

D-stage—The third, or D-stage major equipment is described in the Appendix A.

The D-stage is essentially a conventional chlorine dioxide bleaching stage. Operating parameters at startup and presently are close to those shown in Table C—7. Chesapeake has had a number of mechanical problems with the system, especially in the updraft tube and downflow tower. These problems, their causes and corrective steps are described in the next two sections.

The updraft tube which first receives semi-bleached pulp from the O-stage has undergone a series of most unusual events. It stands in a relatively sheltered spot between the downdraft tower and the bleach plant building. At the top it extends about 3.7 m (12 ft) above the roof level where it joins the top of the downdraft tower.

During high wind conditions in April 1975 the updraft tube broke at the top, fell across a roadway and into the side of the pulp warehouse; the inlet to the downdraft tower was also damaged. This shutdown the bleach plant for twenty days, during which time a completely new updraft tube was fabricated to the same specifications and erected.

From late 1975 to September 1976, numerous patches and repairs were made to the inside surface of the new updraft tube. An attempt failed in September 1976 to reline the entire tube; the new lining would not adhere to the existing substrate, probably because of residual chemical leaching. As a stop gap until a corrective decision could be made, a fine-coat resin lining was applied.

In spite of the high first cost, Chesapeake decided to replace the entire FRP updraft tube with Grade A, 5 mm (3/16-inch) thick titanium. The expected service life is 15 to 20 years versus 1½ to 2 years for FRP. The titanium tube was fitted with three flanged field joints to facilitate field assembly and partial future replacement if required. The new updraft tube was installed in June 1978.

The downdraft tower failed in August, 1975, after approximately 30 months of service. The events surrounding the failure were carefully investigated and fully documented(68). The water drained from the stock in the top section so that a crust formed on the top. Because of the normal amount of water entering the bottom through the agitator seals, the crust was lifted. When the crust pressed against the top conical section, the tower failed at that point rather than the crust. A six-hour outage was required to repair a leak where the new cone section was butted to the existing cylindrical section.

The principal causes of the failure were (a) dewatering of the stock as the level rose in the tower, (b) lack of level alarms at higher levels in the tower and (c) the wall thickness of the tower at the failure points was less than design specifications.

No further major problems have occurred with the downdraft tower. Normal operating level is 16.8 m (55 ft). Alarms sound when the level reaches 18.3 m (60 ft) and 21.4 m (70 ft).

D-stage washer—No more than normal wear problems plagued the D-stage washer except for serious corrosion. How this occurred and what steps were taken to combat and correct the problem are described in the next section.

Anticipating the possibility of corrosion at the D-stage washer all parts normally in contact with the pulp slurry from the D-stage were built of 317 L stainless steel. Even so, corrosion has been, and still is, a major problem with the D-washer.

Not long after the February 1973 startup residual chlorine dioxide levels measured at the D-washer were substantially higher than expected. A system was set up to add sulfur dioxide as an anti-chlor to the bleached pulp slurry leaving the D-tower. Significant corrosion of the washer vat and drum was not observed until September 1974 even though the sulfur dioxide had dropped the vat pH to 2 to 2.5.

After noting the corrosion, the sulfur dioxide addition was stopped and sodium hydroxide added at the dilution pump (for pulp going to the D-washer) to raise the pH to a controlled level of 4.5 minimum. Furthermore, 7 percent caustic was pumped (a) into the space between the washer outer shell and the 317 L vat liner and (b) into the internals of the drum (until half full), both to neutralize any leakage from the vat. Daily checks have been made of residual caustic in these two protective pools. When the caustic residual approach 20 g/L (or 2 percent), new 7 percent solution was pumped into the two spaces.

When it became apparent in early 1977 that even extraordinary maintenance on the vat and drum was no longer feasible a new vat lining and drum of 317 L stainless steel was ordered. Both were installed June 5—9, 1978. Since then, although the pH level has been controlled at 4.5 and above, substantial pitting has occurred. More effective corrective steps are being sought as of December 1978.

Bleached Pulp Quality—

All pulp produced from the D/COD sequence was to be sold on the open market. Therefore, good pulp quality was a necessity. Initially problems in achieving acceptable pulp quality were experienced. These problems were centered in two primary areas. These were first achieving acceptable viscosity and strength levels, and second achieving acceptable shive levels in the final bleached pulp.

Viscosity and strength—Obtaining high pulp viscosities and associated pulp strengths of 300 CSF freeness levels were significant initial problems. These problems existed because our average brownstock viscosity levels were considerably lower than the sample sent to CIL to do their laboratory bleaching work. Also, there was about a 60 percent loss in viscosity across the bleach plant. This loss was occurring primarily across the high temperature D/C stage and the O-stage. The history of pulp viscosity problems was investigated by G. D. Quesinberry and reported to A. D. Fary(12). the following paragraphs outline the problems and solutions.

Chesapeake entered into a contract with CIL to build a bleach plant capable of producing a competitive strength pulp of 88+ GE brightness from the hardwood species native to the Chesapeake procurement areas. Wood species mix would be 50 to 60 percent oak, 20 to 30 percent gum and poplar, 10 to 20 percent other mixed hardwoods, and 5 to 10 percent softwoods. A mill hardwood sample of 21.7 Kappa number (approximately 16 K-No. on 80 cc basis) and 36.5 unbleached viscosity was evaluated by CIL and from this sample they specified that a bleached pulp of 14 to 15 viscosity and 88+ brightness could be produced.

Preliminary testing after start-up indicated that a two-fold problem existed. (1) The brown stock viscosity coming to the bleach plant was considerably lower than the sample submitted to CIL (15 to 25 vs. 36.5). (2) Little information existed as to the optimum conditions for chlorine bleaching of hardwood pulps to maintain viscosity.

A series of laboratory pulp cooks was undertaken to determine the influence of pulp cooking conditions on unbleached pulp viscosity. At this time also, an extensive lab effort was undertaken to devise a more accurate and reproducible technique for running the 0.5 percent CED viscosity test for unbleached samples. A report was issued on the results of this study recommending a reduction in pine content and an increase in K-number to raise unbleached hardwood viscosity. As much as possible these recommendations were followed. The extent of unbleached viscosity improvements was undocumented.

Daily measurements of bleached and unbleached pulp viscosities was begun and the percentage drop through the bleach plant was approximately 60 percent (average bleached viscosity 8.5 vs. 21.3 into the plant) for the approximately 5 months with no indication of operating variables influencing the viscosity drop.

At that time the pulp mill reduced average cooking pressure and began increasing cooking liquor in an effort to increase residual alkali in the black liquor. Two weeks later it was determined that an adequate level of residual chemical had been reached. A report issued later showed no significant difference in viscosity between pulps produced with and without adequate residual chemical.

Laboratory work done over a 2-month period in 1973 showed that chip size can influence the viscosity at a given K-number. Further work awaited the arrival of a chip fractionator to be able to evaluate a practical commercial method of reducing chip size*. Also, lab work showed that there is a considerable difference in unbleached viscosity among wood species, with poplar having a higher viscosity than oak at a given K-number, but the difference in unbleached viscosity is affected by chip size.

In September 1973 urea was used on the D/C stage in the bleach plant to see if any improvement could be made in viscosity out of the D/C stage. No significant change was obtained at a level of (2 lb/ton) and the results were reported.

Lab studies were begun to investigate the factors affecting viscosity drop in a D/C stage. Principal factors investigated were (1) ratio of ClO_2 (2) pH (3) sequential versus blending bleaching liquors.

*Note—This work was initiated but not completed.

Brown stock pH was found somewhat by accident to be the most significant variable, and a mill trial was run. The results were very poor with a loss in viscosity rather than a gain. Further lab work showed that poor mixing intensity was influencing the results obtained in the lab. It was discovered that under high intensity mixing conditions the ClO_2 and Cl_2 bleaching liquors must be blended together and then the reduced brown stock pH gave a high viscosity.

In October and November, 1973 mill trials were run using the blended ClO_2/Cl_2 and adjusted brown stock pH. All tests showed that the D/C drop was reduced to about ¼ of the previous levels. In addition, for 2 to 3 hour periods on two trials, epsom salts ($MgSO_4.7H_2O$) was applied to the oxygen stage pulp. With the combination of reduced brown stock pH (3.0 to 3.5), blended ClO_2 and Cl_2, and magnesium addition in the O-stage (.05 percent Mg on o.d. pulp), the viscosity drop at the O_2 washer was 30 to 40 percent versus the 60 to 65 percent previously observed.)

An extended (24-hour) trial in the bleach plant was run to remove any remaining doubts that the previous trials were not accidents. The results showed that pulp with an acceptable viscosity level had been produced. A report issued in February, 1974 showed that the pulp of higher viscosity levels had superior strength.

Proposed modifications to the bleach plant and screen room were presented to pulp mill production, and in February, 1974 the new portions of the bleach plant system were put on line. The modified system has been in satisfactory operation since then. Bleached pulp viscosity at the dry machine has ranged from 11.0 to 17.2, the average being 12.9 for the first four days of operation, compared to about 8.5 viscosity without pH control and magnesium addition. The standard deviation since the addition of magnesium and pH control has been 1.3.

The significant findings during this study were (1) the negative effect pine content had on unbleached viscosity, and (2) the positive effect an increase in unbleached K-No. had on viscosity. These relationships were developed from lab pulp cooks and are shown in Figures C—16 and C—17. Also very significant were the demonstrated effects of brown stock pH on the viscosity drop through the D/C-stage and the beneficial effect of magnesium in protecting viscosity in the O-stage. The effect of brownstock pH on D/C stage viscosity drop is shown in Figure C—18. The effect of the addition of magnesium to the O-stage was to reduce the percentage loss of initial incoming brownstock viscosity by about 7 percent or 2 viscosity units.

By proper understanding and use of the solutions obtained during these initial viscosity problems, we have been able to maintain a competitive market pulp viscosity level. Also our bleached pulp physical strength levels (burst, breaking length, and tear), when measured at the same viscosity and pine content have compared favorably with competitive pulps at freeness levels below about 400 CSF. At freeness levels greater than 400 CSF burst and breaking lengths are lower than competitive pulp (see Figures C—19 and C—20). A recent report(13) concluded that the lower strength at the high freeness levels was not due to any detrimental effect of the oxygen bleaching process. Rather it was due to the lower strength level of our unbleached pulp.

Shive and dirt levels—Shive levels in the final bleached pulp became a serious problem early in 1975 when the bleached hardwood pulp market became very tight. In examining the shives contained in the bleached pulp it was found that over 80 percent were pine. An extended study was undertaken to determine the source of these pine shives. The findings of this study were included in an August 12, 1975 memo regarding "Pine Contamination in Bleached Hardwood Pulp"(14).

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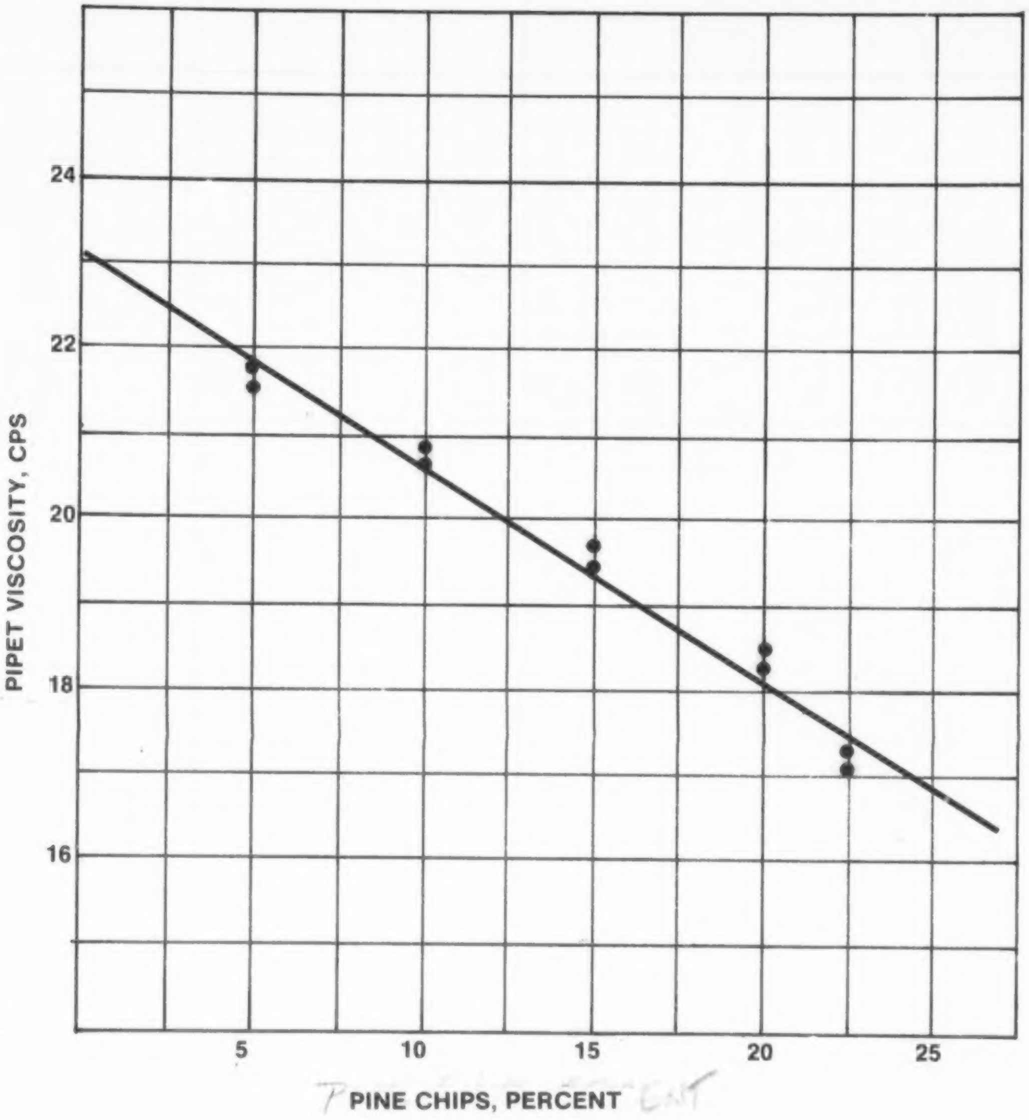

Figure C—16. Viscosity vs. pine content.

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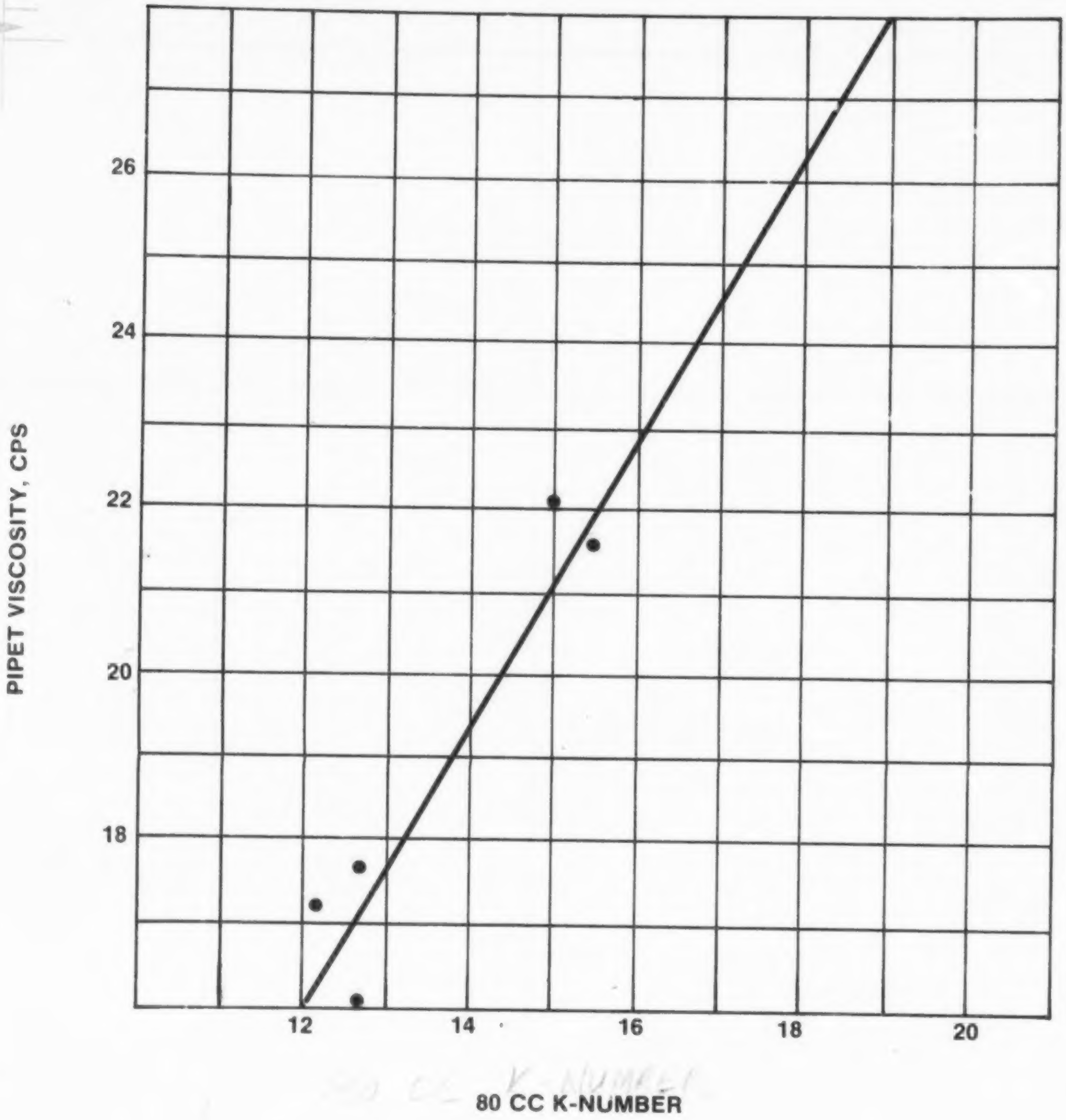

Figure C—17. Viscosity vs. K-number.

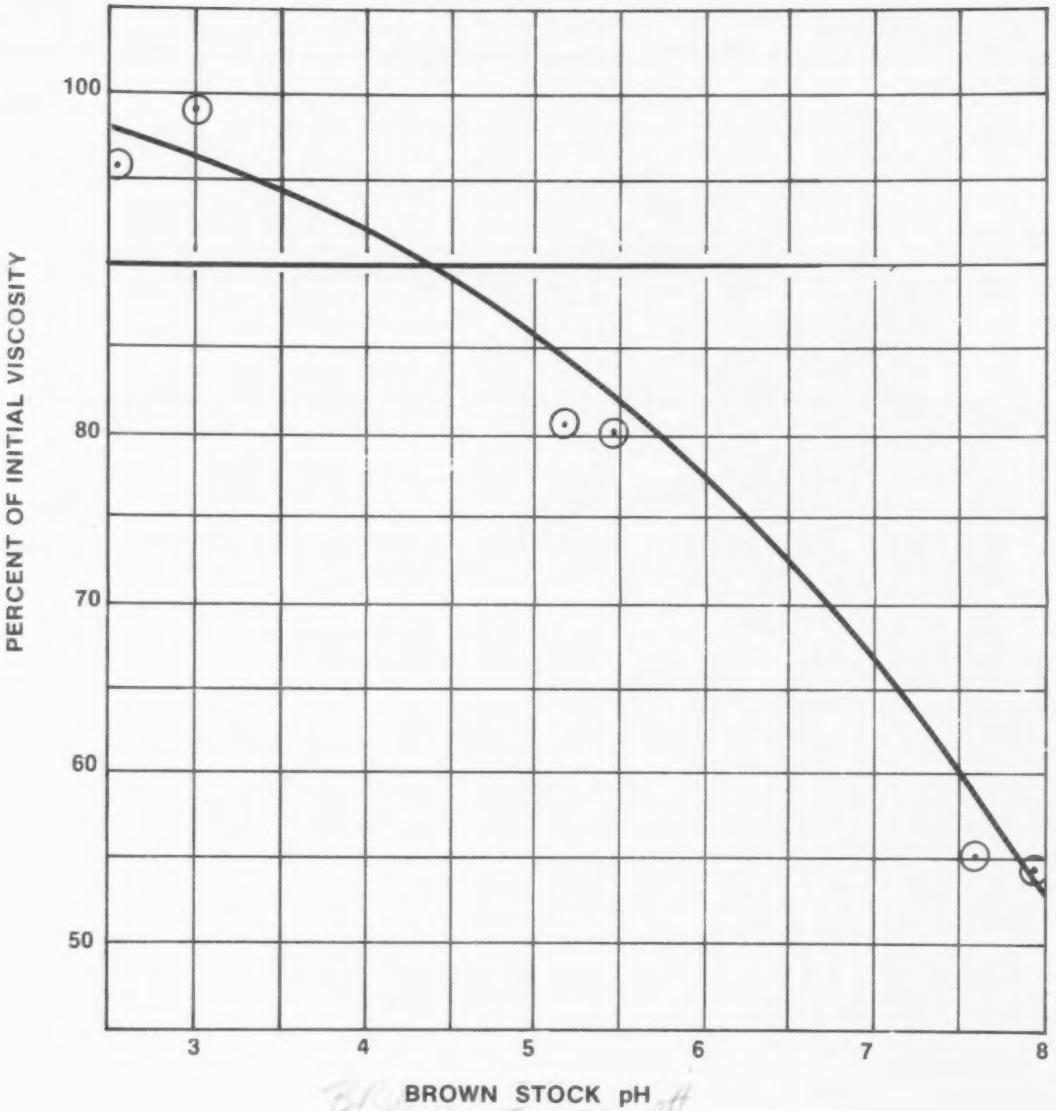

Figure C—18. Viscosity vs. brownstock pH.

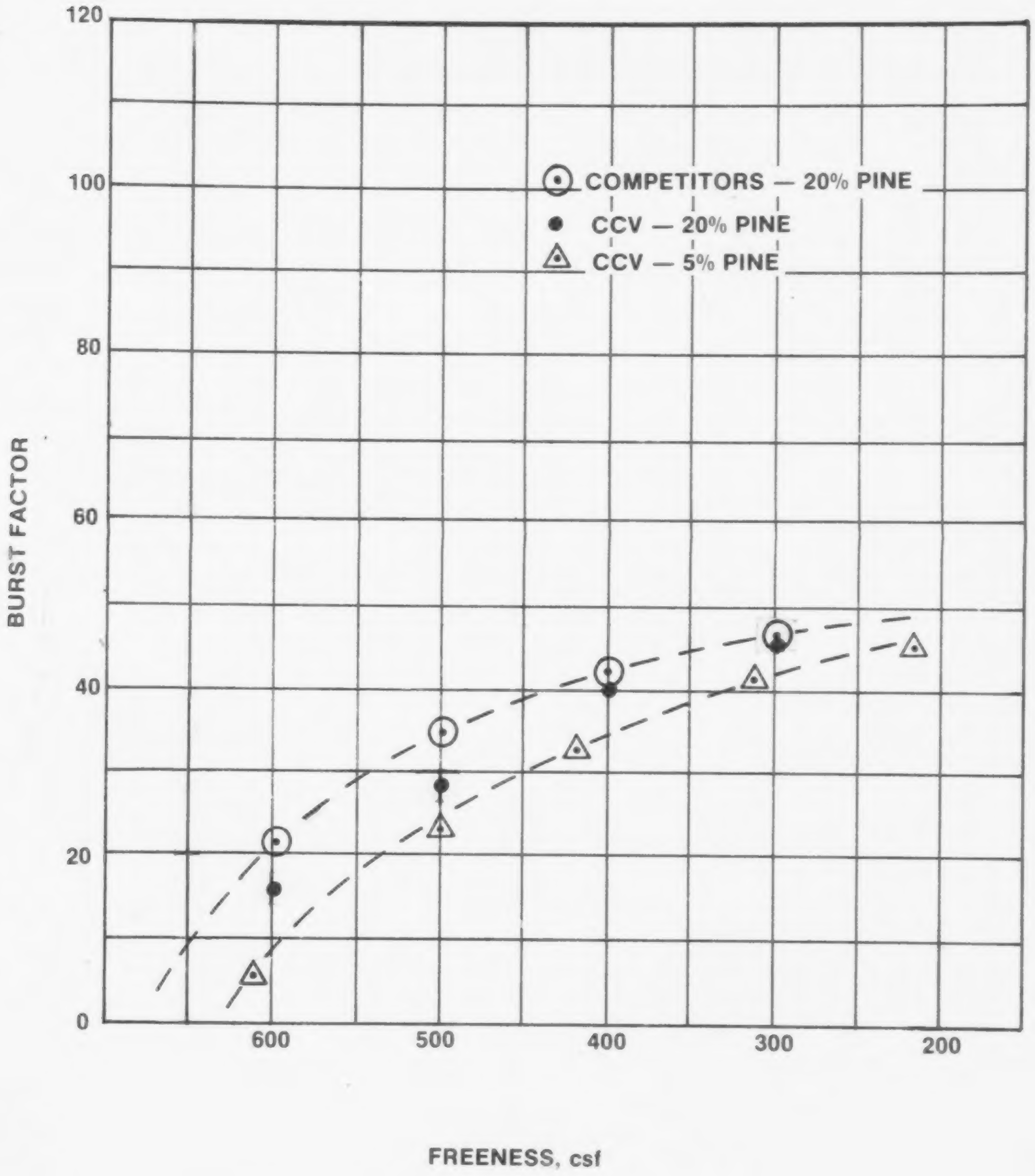

Figure C—19. Burst vs. freeness for CCV and competitive pulps.

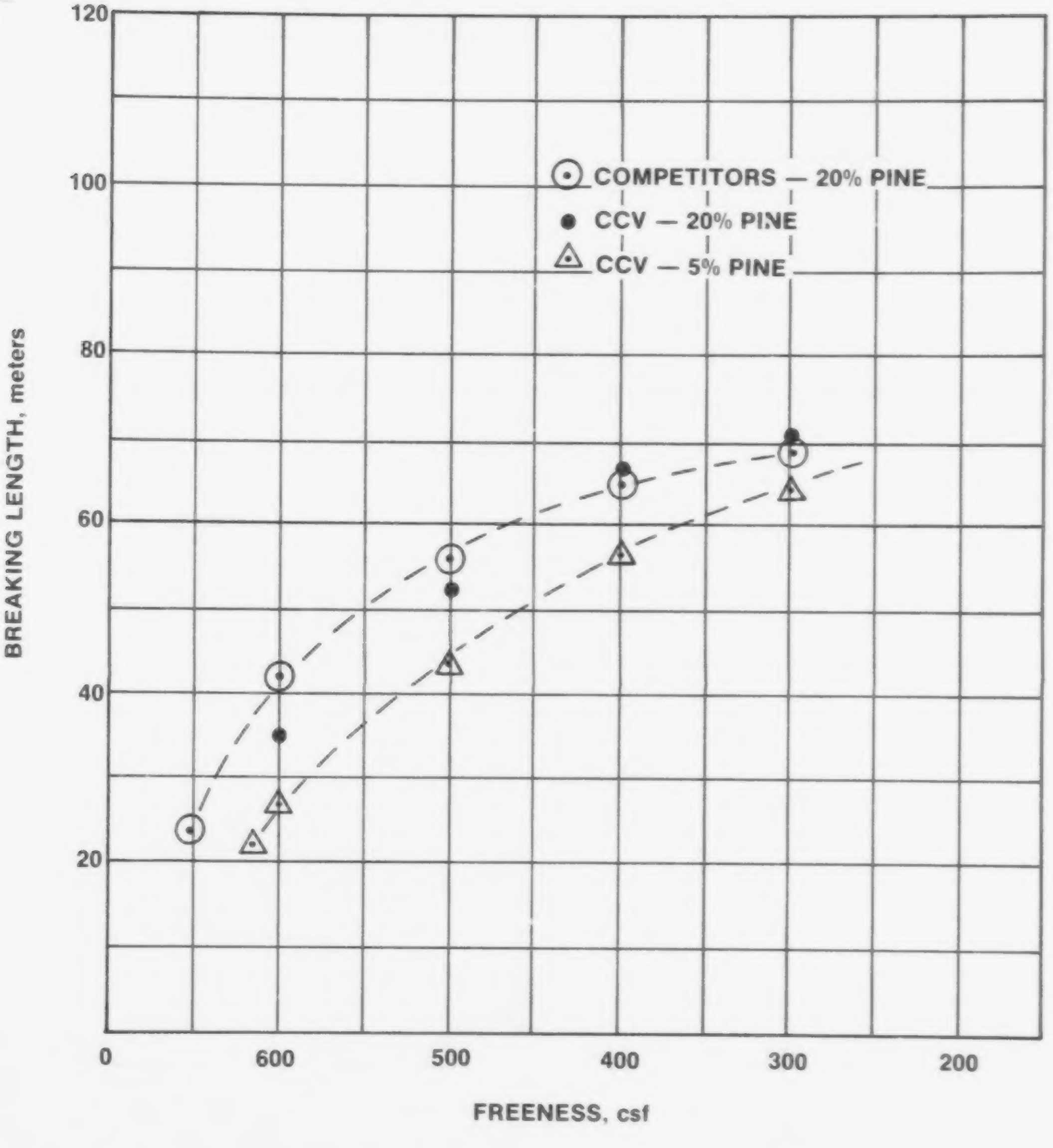

Figure C—20. Breaking length vs. freeness for CCV and competitive pulps.

This report showed that during high shive levels on the dry machine, over 80 percent of the shives present in the bleached hardwood pulp have been identified as pine. In an attempt to determine the source of these shives, samples of unbleached hardwood pulp were taken from the discharge of the batch vacuum washers during two consecutive hardwood runs and analysed for hardwood content. The purpose of these samples was:

(1) To determine the extent of mixing of hardwood and pine chips prior to the batch digesters.

(2) To determine the extent of mixing of hardwood and pine pulps on the batch vacuum washers during transitions from pine to hardwood and vice versa. Also to determine if these transitions can be a source of pine shive problems in the bleached pulp.

(3) To determine the length of time needed to carry-out transitions from pine to hardwood on the batch vacuum washers so that problems of pine contamination of the hardwood to the bleached plant, and hardwood contamination of the pine to the paper mill will be minimized.

Conclusions drawn from the data collected are as follows:

(1) The mixing of pine and hardwood ahead of the batch digester is substantial and variable. The hardwood content of the pulp from the batch vacuum washers, over two consecutive hardwood runs, averaged 91 percent with a range of values from 84 to 96 percent. The effect of any hardwood variation due to transitions is not included in the indicated range of hardwood values.

(2) The mixing of pine and hardwood did occur at high levels for short periods of time (5 to 10 min) during transitions on the batch washers. During the four transitions measured, the pine content of the pulp when routed to the bleach plant brown stock storage was 14 percent 63 percent, and 93 percent. The hardwood content of the pulp when routed to the paper mill brown stock storage tank (no. 4 stock tank) was 11 percent, 27 percent, 1 percent, and 4 percent. The highest hardwood content (27 percent) occurred when a reject digester was the first digester blown after the transition from hardwood to pine.

(3) The stock retention time between the blow tanks and the batch washer discharge is between 5 and 10 min. Therefore, the time required to go from a high percentage of hardwood (85 percent) to a high percentage of pine (85 percent) on the batch washers normally takes less than 10 min. However, these transitions can be longer if a reject digester, or No. 1 digester, is the first digester washed after a transition occurs.

(4) No. 1 digester is a source of pine contamination to the bleach plant and hardwood contamination to the paper mill. The bin above this digester collects chips which are carried over when loading other digesters.

In addition to the above, the data collected allows us to draw a preliminary conclusion that transitions on the batch washer may cause shive problems on the dry machine. Dry machine shive problems did occur at the predicted time that transitions high in pine content would have reached the dry machine. These predicted times were calculated assuming plug flow through the brown stock storage, bleach plant and bleach plant storage.

To minimize mixing of pine and hardwood the following recommendations are made:

(1) Operating guidelines should specify switching into a transition-grade stock tank within 5 min after switching between pine or hardwood blow tanks. Stock should continue into the transition stock tank for a minimum of 5 min. It may be necessary to continue putting pulp into the transition tank for time periods longer than 5 min to give the operator time to obtain good operation of the hot stock screens and refiners. Also a longer time will be needed if a reject digester or no. 1 digester is the first digester blown.

(2) Additional samples at shorter time intervals should be collected off the batch washers to better determine the time needed for a transition to occur.

(3) Further consideration should be given to the problem of pine contamination from no. 1 digester.

(4) Every effort should be made to control mixing of hardwood and pine chips in the woodyard.

Results from this study showed we were getting mixing of pine and hardwood chips ahead of the batch digesters. Also we were getting pulp mixing when we switched from hardwood to pine on our brownstock washers. Mixing of chips and pulp will occur in a system like Chesapeake's in which pine and hardwood must be handled, up to brownstock storage, through the same process equipment. A simplified flowsheet of our pulping system is shown in Figure C—21.

It was understandable that mixing of hardwood and pine pulp during transitions on the brownstock washers would cause a problem. The pine pulp was being used for making unbleached linerboard and was being cooked to about a 40 K-No. target. Normal brown stock K-No. levels for bleachable grade pine pulp are about 20 K-No. Shive problems could be expected when bleaching pine at 40 K-No. This problem of mixing of pine and hardwood was solved by following proper time schedules during pulp washer transitions.

Understanding the effect of the mixing of pine and hardwood chips during the pulping process was more difficult and required additional laboratory work. Laboratory digester cooks on mixtures of pine and hardwood showed that pine is not well cooked and therefore, will have a high K-No.(Figure C—22). For example, the pine K-No. for a 90 percent hardwood and 10 percent pine mixture will be in the range of 25 to 48 for a composite hardwood K-No. range of 12 to 19. These high pine K-Nos. would be more difficult to bleach and could cause shive problems. This is particularly true in our case since the hardwood stock for our bleach plant is hot stock refined and screened with no rejects being removed from the brown stock system. Therefore, none of the high K-No. pine will be removed prior to the bleach plant (Figure C—21). We are continuing to use this flowsheet for processing our hardwood pulp prior to bleaching. In order to minimize the problems this flowsheet presents we are taking all steps possible to prevent mixing of pine and hardwood chips ahead of the digester.

It should be noted that it is very unusual for a closed brown stock system utilizing hot stock refining and screening to be used in preparing bleachable grade hardwood pulp. Normally, there is a concentrated brown stock stream rejected from knotters, screens, centrifugal cleaners, or combinations of these cleaning devices. This reject stream is normally recooked or used in a product requiring less cleanliness. In Chesapeake's case there is no reject stream and all rejects are refined and sent to the bleach plant. Also it is normal in a bleached mill to have chip screens to remove oversize chips as well as fines. Chesapeake has no chip screens. The ability of the oxyen bleach system to produce a pulp which is low in dirt and shive levels from a 16 K-No. pulp processed by Chesapeake's flowsheet shows the large dirt and shive removal potential of the oxygen bleach plant.

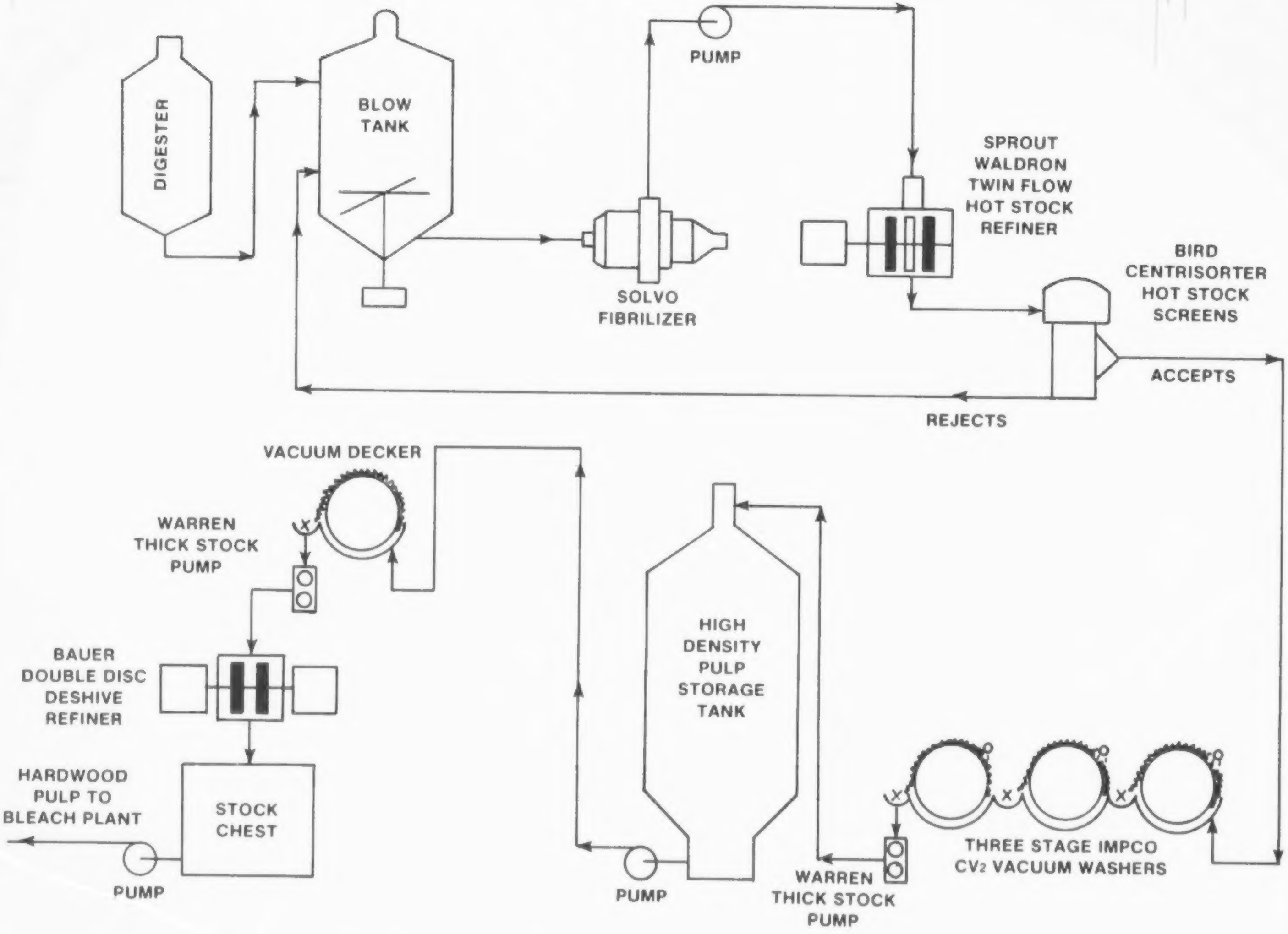
DIGESTER
BLOW TANK
SOLVO FIBRILIZER
PUMP
SPROUT WALDRON TWIN FLOW HOT STOCK REFINER
BIRD CENTRISORTER HOT STOCK SCREENS
ACCEPTS
REJECTS
VACUUM DECKER
WARREN THICK STOCK PUMP
BAUER DOUBLE DISC DESHIVE REFINER
HARDWOOD PULP TO BLEACH PLANT
STOCK CHEST
PUMP
PUMP
HIGH DENSITY PULP STORAGE TANK
WARREN THICK STOCK PUMP
THREE STAGE IMPCO CV2 VACUUM WASHERS

Figure C—21. Brownstock flowsheet for bleachable grade hardwood pulp.

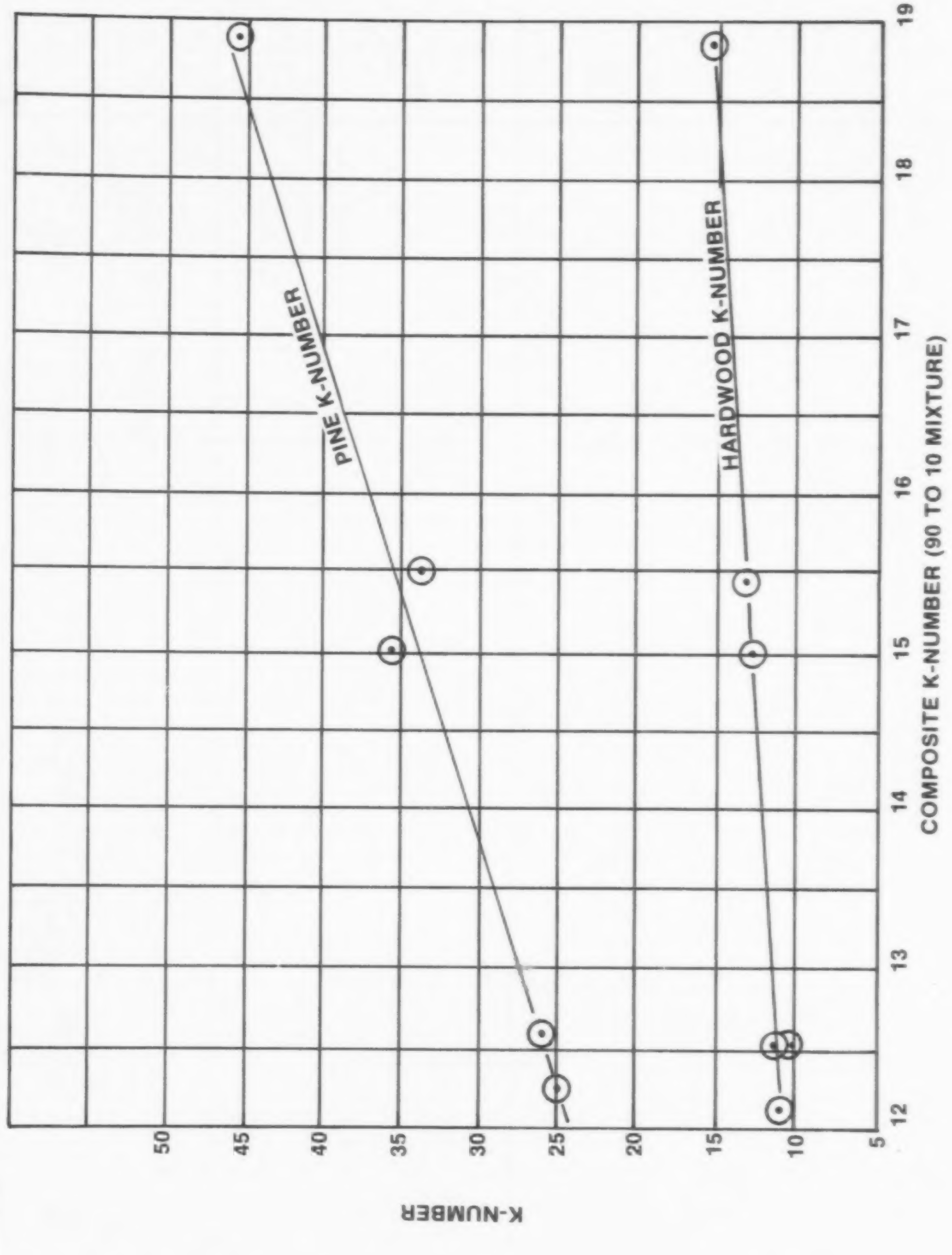

Figure C—22. K-number of pine and hardwood mixtures (90 percent to 10 percent mixture of hardwood and pine by weight).

Other important bleach pulp properties—Minimum problems were experienced in obtaining acceptable bleached brightness or dirt levels at design tonnage rates. Also bulk and opacity which are very important properties to be maintained in bleached hardwood pulps have been at or above acceptable levels. Table C—15 compares the properties of pulp from the D/COD sequnce to that of five Southern hardwood pulps. The comparison is made at 10 to 10.5 viscosity and a 300 CSF freeness level. Also data is given at two different levels of pine (5 and 20 percent) since pine content can have a large effect on the properties of the final pulp. As can be seen from this Table C-15, properties of the oxygen bleached pulp are very similar to those of the pulps bleached by conventional sequences.

Bleach Plant Effluent By Stage—

Development of stage by stage data for flow, BOD_5, color, sodium and chlorides was necessary to determine the pollution reduction potential of the D/COD sequence with and without recycling of O-stage effluent. Original projections of the quantity of color, BOD_2, and chlorides from the D/COD sequence were made by CIL based on laboratory studies(15). These data are included in Table C—16 and indicate combined bleached effluent BOD_2 and color levels of some 16 and 52.5 kg/metric a.d. ton (32 and 105 lb/a.d. ton). Also, laboratory data indicated the oxygen stage contained approximately 60 percent of the BOD_2 and 50 percent of the color in a concentrated stream of about 4507 L/metric a.d. ton (1080 gal/ a.d. ton). These data were developed by the laboratory bleaching of a representative sample of Chesapeake hardwood pulp (16 K-No. to 88 brightness).

Following startup of the bleach plant, data were collected on a routine basis to measure properties of the combined effluent. Data collected between 2/15 and 4/15/74 indicated that combined effluent BOD_5 and color levels were 14.5 and 42.5 kg/metric a.d. ton (29 and 85 lb/a.d. ton) respectively (Table C—17). Also, initial stage-by-stage data collected in July and August 1974 indicated combined effluent BOD_5 and color levels were 11.5 and 42 kg/metric a.d. ton (23 and 84 lb/a.d. ton). Both of these sets of data agreed reasonably well with each other and with the initial laboratory data from CIL. Therefore, these results were initially reported as typical of those to be expected from a D/COD sequence(16). A 23 percent reduction in BOD_5 and a 58 percent reduction in color were reported using 17.5 and 100 kg/metric a.d. ton (35 lb/a.d. ton and 200 lb/a.d. ton) as typical BOD_5 and color contributions from conventional sequences.

Following startup and operation of the bleach plant sump and the waste treatment plant, it was apparent that the bleach plant effluent loads were higher than those previously reported. Therefore, an additional study of stage-by-stage and combined effluent data was undertaken in July 1976 and reported in a November 3, 1976 memo from G. D. Quesinberry to G. R. Howell(17).

This memo reported on the question of why bleach plant effluent loadings were higher than previously reported. In an effort to answer that question, a bleach plant effluent material balance was attempted. The balance was made by taking daily composite samples of each of the seal tank overflows combined effluent for one week. Various components of the composite samples were analyzed. Using a computer technique for solving simultaneous equations, flows for each stage were estimated. The bleach plant effluent flowmeter was also checked using the lithium salt tracer method. Table C—18 shows a summary of the effluent material balance data for the trial period.

The large difference between the effluent flows from 1974 and 1976 cannot be readily explained. In an effort to account for the present flow, the information in Table C—19 was compiled for a 254 metric a.d. tons/day rate (280 a.d. tons/day).

Although these numbers are only approximate, the total of 91.5 L/s (1450 gpm) is within 5 percent of the value from the material balance data.

TABLE C—15. PROPERTIES OF OXYGEN AND CONVENTIONALLY BLEACHED PULPS AT 300 CSF

Mill	Burst (factor)	Breaking length (m)	Tear (factor)	Bulk cm^3/g	Opacity	Brightness units (GE)	Shives (number)	Dirt (cm^2/m^2)	0.5% CEP viscosity (centipoise)	Hard-wood (%)
A	43	6800	57	1.47	74.3	85.6*	—	—	10.0	95
B	46	6900	55	1.46	75.0	86.1*	0	0.2	10.3	90
C	51	7050	95	1.47	—	87.0*	1—2	0.6	10.4	85
D	47	7000	95	1.47	71.0	84.2*	5—6	1.1	11.1	70
E	49	7200	88	1.47	75.8	84.3*	—	—	10.5	61
F	—	—	—	—	—		1—4	0.7	—	—
G	—	—	—	—	—		7—8	0.7	—	—
H	—	—	—	—	—		1—2	1.6	—	—
Mean ±	47	6990	78	1.47	74.0	89.0	2½—3½	0.8	10.3	87±
	± 3	± 150	± 20	± .005	± 2.1			±.5	±.2	19
Chesapeake pulp	43	6600	85	1.49	75.5	88+	Less than 3	Less than 1	10.5	95
CCV pulp with pine added	47½	6675	90	1.50	72.7	88+	Less than 3	Less than 1	11.0	80

*Samples had naturally aged. Estimated original brightness several units higher.

Since we have not added any significant source of water to the bleach plant process stream since 1974, it would appear that the total effluent in 1974 reported as 66.3 L/s (1050 gpm) is low. The measured value of 107 L/s (1700 gpm) for the present data would appear to be more accurate. The 1974 data was based on measurements taken from a Parshall flume. The 1976 data was based on orifice measurements with calibration by a lithium salt trace measurement.

Ratioing the 1974 data by the ratios of the flows (1715/1052) gives 21.3, 62.5, and 42.3 lb/metric a.d. ton (42.5, 125, and 84.6 lb/ad.d. ton) of BOD_5, color and chlorides for that period. This would seem to indicate that the present bleach plant effluent load is about the same as it was previously, and that the previously reported flows were in error on the low side.

TABLE C—16. BLEACH PLANT EFFLUENT STAGE-BY-STAGE LEVELS CALACULATED FROM CIL LABORATORY DATA

Source	Flow L/a.d. metric ton (gal/a.d. ton)	BOD_5	Color	Chloride
		All values in kg/a.d. metric ton (lb/a.d. ton)		
D/C	————	3.95 (7.9)	24.8 (49.5)	19.7 (39.3)
O_2	4507 (1080)	9.5 (19.0)	26.8 (53.6)	2.1 (4.2)
D		2.55 (5.1)	1.05 (2.1)	3.5 (7.0)
Total	————	16.0 (32.0)	56.2 (105.2)	25.1 (50.2)

TABLE C—17. BLEACH PLANT EFFLUENT STAGE-BY-STAGE LEVELS JULY AND AUGUST, 1974

	Flow L/s (gpm)	BOD_5	Color	Sodium	Chloride
		All values in kg/a.d. metric ton (lb/a.d. ton)			
D/C	36.2 (575)	4.55 (9.1)	15.4 (30.8)	7.9 (15.8)	17.45 (34.9)
O_2 O_2	13.4 (213)	5.8 (11.6)	25.55 (51.1)	12.0 (23.9)	5.4 (10.8)
D	9.6 (152)	1.25 (2.5)	1.1 (2.2)	2.65 (5.3)	3.0 (6.0)
Calc. total	59.2 (940)	11.6 (23.2)	42.05 (84.1)	22.5 (45.0)	25.9 (51.7)
Combined effl.*	72.6 (1052)	14.5 (29)	42.5 (85)	———	3.0 (5.9)

*As of 2/15—4/14/74

It is clear from the study in July 1976, that our initial reported information on BOD_5 and color were in error and that BOD_5 levels of 24 kg/metric a.d. ton (48 lb/a.d. ton) and color levels of 70.5 kg/metric a.d. ton (141 lb/a.d. ton) were representative of the combined effluent from the D/COD sequence. The data further showed that the O-stage contributed about 54 percent of the BOD_5 and 45 percent of the color. This was in good agreement with the previous stage by stage study.

Table C—20 compares the color, chloride, and BOD_5 levels of the D/COD sequence with and without O-stage recycle, to that of conventional bleach plants. Data on conventional sequences is based on a 1974-1975 survey of 23 mills by NCASI(18). Data on chlorides is taken from literature references and available NCASI mill data(7, 18, 19). It is clear that without O-stage effluent recycle, the

TABLE C—18. D/COD STAGE-BY-STAGE EFFLUENT DATA

	Flow		BOD_5		Color		Sodium		Chlorides	
	L/s (gpm)		All values in kg/metric a.d. ton (lb/a.d. ton)							
D/C	49.1	(780)	7.65	(15.3)	34.5	(69.0)	9.2	(18.4)	26.5	(53.0)
O	19.2	(305)	12.8	(25.6)	32.0	(64.0)	12.8	(25.5)	5.3	(10.5)
D	24.6	(390)	3.35	(6.7)	3.95	(7.9)	5.7	(11.4)	5.3	(10.3)
Process water calculated	15.1	(240)	0	(0)	0	(0)	0.95	(1.9)	0.35	(0.7)
Total	108	(1715)	23.8	(47.6)	70.5	(140.9)	28.6	(57.2)	37.3	(74.5)
O-stage per-cent of total)	1.12	(17.8)	26.9	(53.7)	22.7	(45.4)	22.3	(44.6)	7.05	(14.1)

TABLE C—19. EFFLUENT SOURCES BY BLEACHING STAGE

Source	D/C stage L/s (gpm)		O_2 stage L/s (gpm)		D stage L/s (gpm)	
Consistency change*	25.2	(400)	-9.5	(-150)	0	(0)
Showers	9.5	(150)	22.0	(350)	15.8	(250)
Chemicals	9.5	(150)	1.6	(25)	4.7	(75)
Wire cleaning	3.2	(50)	3.2	(50)	3.2	(50)
Condensed steam	———		1.6	(25)	1.6	(25)
Total	47.3	(750)	18.9	(300)	25.2	(400)

*The consistency change between the stock coming into the stage compared to the stock leaving that stage.

D/COD sequence does not result in any BOD_5 reduction. However, about 30 percent less color is created. With O-stage recycle, assuming complete destruction of BOD_5 and color, a 40 percent reduction in BOD_5 and a 65 percent reduction in color could be realized.

It should also be noted, as discussed in "Bleach Pulp Quality" that the D/COD sequence BOD_5 and color levels in Table C—20 result from bleaching hardwood pulps at about 16 K-No. It has been our experience that brownstock K-No's. for conventional bleach plant sequences are normally in the 10 to 12 range. Lower color and BOD_5 and chloride levels would be expected at these lower unbleached K-No's.

Recycling O-Stage Effluent—
The building and operation of a pilot plant to reduce chlorides in white liquor so as to maximize return of the oxygen stage effluent to chemical recovery was an original objective of the EPA grant. As the overall project progressed, however, it became clear to Chesapeake that the completion of this objective within the grant time period would be most difficult and also questionable in light of progressing technical developments. The large amount of time required to solve initial operating problems with the bleach plant, O_2 plant, WWTP and BLO systems severely taxed all available technical manpower and left little, if any time to consider work on the pilot crystallizer until late in the grant period. By that time two significant technical advances had taken place. First, CIL had completed sufficient work on their OD/CED sequence to be convinced it was a viable bleaching process. A letter from CIL dated September 5, 1974 stated that they had continued development work on oxygen bleaching and were prepared to offer technology associated with bleaching high yield—55 to 60 percent pulp(20). This letter suggested an OD/CED sequence if Chesapeake expanded its bleach plant. This sequence would allow recycle of O-stage effluent and significant BOD_5 and color reduction without an increase in recovery cycle chloride levels. Therefore, no crystallizer would be

TABLE C—20. COMPARISON OF MILL EFFLUENTS FROM D/COD AND CONVENTIONAL SEQUENCES

Sequence	Flow L/s (gpm)	BOD_5	Color	Chlorides
		All values in kg/metric a.d. ton (lb/a.d. ton)		
D/COD	554 (8,800)	23.8 (47.6)	75 (149.9)	74.5 (37.25)
D/COD with O-stage recycle	———	11.0 (22.0)	38.5 (76.9)	———
Conventional 4 to 6 stage hardwood bleach sequences	1,241—1,764 (19,700--28,000)	17—20.5 (34—41)	105—117.5 (210—235)	29—42.5 (58—85)
Percent reduction from conventional sequence levels with O-stage recycle	———	17.5—23 35—46	31.5—33.5 (63—67)	———

necessary. Also, by mid-1973, Erco Envirotech Ltd. felt that their evaporation-crystallization salt recovery process, conceived initially by Reeves and Rapson, was ready for commercial installation(21).

In light of these developments, it did not appear that efforts by Chesapeake to develop a crystallizer in conjunction with an oxygen bleaching system was necessary. However, sufficient data was collected to determine the chloride concentration in white liquor which would result from recycling the O-stage effluent from a D/COD sequence to our recovery system. This is explained below.

O—stage effluent chloride levels—O-stage effluent chloride levels were determined in two different studies. The first set of data was collected in 1974 before installation of the D/C stage washer. This data, and the associated problems with this data, have been discussed in the previous section. A second set of data was collected in 1976 after the D/C washer installation. This data was summarized in Table C—18 and showed a level of 5.3 kg Cl^- metric a.d. ton (10.5 lb Cl^-/a.d. ton) from the O-stage. It should be noted at this point that the procedure used to determine chlorides measured only the free Cl^- ion. It did not measure origanically bound Cl^- (see Appendix B).

Chloride levels in white liquor—The chloride levels in white liquor by recycling the O-stage effluent can be calculated easily is a simplifying assumption is made. This assumption is that there is no differential losses of sodium or chlorides from the recovery system. Therefore, the chloride/sodium ratio (Cl^-/Na^+) which will be reached at equilibrium in the mill white water will be the same as the Cl^-/Na^+ ratio in the chemical make-up to the recovery system. This assumption was checked by a mill sodium and chloride balance which was done for the years 1975 and 1976(69). In this report levels of chlorides in white liquor were calculated from the ratio of Cl^- to Na^+ in our chemical makeups and compared to actual measurements of chloride measurements in white liquor. Results showed calculated levels of 4.6 and 6.1 g/L (28.6 and 38.1 lb NaCl/100 ft^3). Corresponding measurements of actual chloride levels in white liquor were 5.0 and 6.2 g/L (31.5 and 38.5 lb NaCl/100 ft^3). A calculation of the change in white liquor chlorides from recycling O-stage effluent is shown below using appropriate data on sodium and chloride make-up, pulp tonnages, etc.

Given: Production data for year 1976

1.	Unbleached pulp production	952.5 metric tons/day (1050 tons/day)
2.	Bleached pulp production	74,640 metric tons/year (82,277 tons/year)
3.	Cl^- from O-stage (bleached)	5.85 kg/metric ton (11.7 lb/ton)
4.	Na^+ from O-stage (bleached)	14.1 kg/metric ton (28.3 lb/ton)
5.	Na^+ make-up (unbleached)	19.3 kg/metric ton (38.5 lb/ton)
6.	Cl^- make-up (unbleached)	0.75 kg/metric ton (1.50 lb/ton)
7.	Active alkali (unbleached)	2.95 kg/metric ton (5.90 lb/ton)
8.	Active alkali in white liquor	106.2 g/L (6.63 lb/ft^3) 95.0 g/L Na^+ (594 lb/100 ft^3 Na^+)
9.	Cl^- in NaOH make-up	0.87 percent
10.	Differential loss of Na^+ or Cl^- in recovery system	none

To Find: 1976 equilibrium Cl^- level if O-stage flitrate is recycled

Calculations:

1. Amount of Cl^- per unit of unbleached pulp

$$\frac{74{,}640 \text{ metric tons/year} \times 5.85 \text{ kg } Cl^-\text{/metric ton}}{952.5 \text{ metric tons/day} \times 350 \text{ days/year}} = 1.31 \text{ kg } Cl^- \text{ metric ton}$$

(2.62 lb Cl^-/ unbleached ton)

2. Reduction in NaOH make-up

$$\frac{74{,}640 \ \frac{\text{metric}}{\text{tons/year}} \times \left(14.1 \ \frac{\text{kg } Na^+/}{\text{metric ton}} - 5.85 \ \frac{\text{kg } Cl^-/}{\text{metric ton}} \times \frac{23}{35.5}\right)}{952.5 \text{ metric tons/year} \times 350 \text{ days/year}} = 2.31 \text{ kg NaOH/ metric ton}$$

(4.62 lb/ton)

2.31 kg NaOH/metric ton x 0.87 % Cl^- = 0.02 kg Cl^-/metric ton (0.04 lb/ton)

3. Chloride in chemical make-up, kg/metric ton (lb/ton)

Present make-up		Cl^- in O_2 effluent		Reduction in caustic make-up		Chloride in chemical make-up
0.75 (1.50)	+	1.31 (2.62)	−	0.02 (0.04)	=	2.04 (4.08)

4. White liquor concentration of Na Cl with O-stage recycle

$$\frac{2.04 \text{ kg } Cl^-\text{/metric ton}}{19.3 \text{ kg } Na^+\text{/metric ton}} = \frac{x}{95.0 \text{ g/L } Na^+}$$

$x = 10.0$ g/L (62.8 lb/100 ft³)

$$10.0 \text{ g/L} \times \frac{58.5}{35.5} = 16.5 \text{ g/L NaCl} \quad (103 \text{ lb NaCl/100 ft}^3)$$

5. White liquor concentration of Na Cl without O-stage recycle

$$16.5 \text{ g/L} \times \frac{0.75 \text{ kg/metric ton } Cl^-}{2.04 \text{ kg/metric ton } Cl^- \text{ in make-up}} = 6.1 \text{ g/L}$$

As the above calculations show, the recycle of O-stage effluent would increase our white liquor NaCl levels from about 6 to about 17 g/L. It is not within the scope of this report to document the probable effects of this increase in white liquor chloride levels on total mill operations. A great deal of attention is presently focused on this problem by the design and startup of the "Rapson-Reeves" closed mill concept at Thunderbay. Literature on this problem has been reviewed by Douglas Reeve(22) in conjunction with the development of this concept. It is interesting, however, to point out one of the apparent design criteria of this salt recovery system installed at Thunderbay. This is the indicated equilibrium white liquor salt concentration of 12 g/L in the closed cycle mill with salt removal by crystallization. If this is in fact the design salt concentration level, it is unlikely that the development of a crystallization process to lower salt levels from 17 to 12 g/L have been justified in Chesapeake's case.

The still overriding question to answer, however, is what effect elevated chloride levels from recycling bleach plant effluents will have on overall pulp mill operations. This very complicated question will be one of the most important facing the industry in the near future.

Bleach Plant Chemical Usage

Typical usage of chemicals in the bleach plant is summarized in Table C—21. Data on total chemical usage is the average for a four-week period in July and August 1976 (7/11 to 8/8/76). This includes the time period in which the stage-by-stage data on effluent BOD_5, color, chlorides, etc., was taken. Average unbleached K-No. for this four-week period was 15.7, bleach average brightness 90 to 90.5, and average tonnage 246 metric a.d. tons/day (271 a.d. tons/day). This data can be used to determine the chemical cost of the D/COD bleach sequence.

TABLE C—21. D/COD BLEACH CHEMICAL USAGE

Bleach chemical	D/C stage	O stage	D stage	pH control D/C and D-stage	Total
Chlorine, percent	3.8	—	—	—	3.8
Chlorine dioxide, percent	.75	—	1.0	0	1.75
Sodium hydroxide, percent	—	3.0	0	1.6	4.6
Oxygen, percent	—	1.7	—	—	1.7

AIR SEPARATION PLANT

A detailed description of the major oxygen plant operating equipment is presented in Appendix A. This section of the report: (a) outlines the various steps taken by Chesapeake to select the size and type of oxygen plant, (b) describes the process, (c) discusses the plant startup including operating problems, (d) summarizes plant production, and (e) details plant operating costs.

Selecting Size and Type of Oxygen Plant

The fundamental concerns in selecting the oxygen plant were: (a) size of plant, (b) oxygen purity, (c) total installed cost, (d) operating costs, and (e) plant reliability. The oxygen requirements of the bleach plant were initially economically evaluated based on the purchase of liquid oxygen. An on-site 34.1 m^3 (9,000 gallon) insulated liquid oxygen storage tank, complete with an ambient vaporizer capable of delivering 4.5 to 6.4 metric tons/day (5 to 7 tons/day) of gaseous oxygen at 1034 kPa gage (150 psig) was included in the evaluation.

Additional environmental uses of elemental oxygen, including oxygen activated sludge waste treatment and weak black liquor oxidation, were independently evaluated. Table C—22 summarizes the estimated capital and operating costs of the investigated oxygen plant alternatives.

After studying the various alternatives, the adsorption type of oxygen plant was eliminated from consideration due to: (a) higher capital and operating costs (at the production rates investigated), (b) oxygen purity required by the bleach plant (99.5 percent) and (c) decreased utilization of oxygen by the waste treatment plant (due to lower oxygen purity). In addition, the purchase of liquid oxygen for bleach plant operation and as backup for peak oxygen demands and/or plant outages, weighed heavily against the adsorption type plant. Therefore, a cryogenic oxygen plant with a product purity of 99.5 percent was selected as the basis for design.

The economic evaluation justified the use of elemental oxygen for the bleach plant, waste treatment plant and weak black liquor oxidation system. Initially, a 77.1 metric tons/day (85 tons/day) oxygen plant was investigated to supply the total oxygen requirements of these systems. However, at this point, Chesapeake's president suggested that the oxygen plant capacity be increased to 90.7 metric tons/day (100 tons/day) on his assumption that additional uses for oxygen would be found in the manufacturing process. This assumption was proven correct with the installation of a system to add oxygen to the lime kiln to maintain production while reducing the kiln TRS emissions.

As a result of these studies, a 90.7 metric tons/day (100 tons/day) cryogenic oxygen plant designed to produce 81.6 metric tons/day (90 tons/day) of gaseous and 9.1 metric tons/day (10 tons/day) of liquid oxygen at a product purity of 99.5 percent was purchased from Burdett Oxygen Company of Norristown, Pennsylvania. The plant was supplied by Superior Air Products of Sayreville, New Jersey, and installed under Superior's supervision by The Rust Engineering Company, of Birmingham, Alabama.

A purchase-sale agreement was signed by Chesapeake Corporation and Burdett Oxygen Company whereby Chesapeake sold excess liquid oxygen to Burdett with maximum sales limitations of 26,055 standard m^3 (920,000 scf) per day and 246,385 standard m^3 (8.7 million scf) per month.(7) The agreement also stipulated that Chesapeake would purchase required liquid oxygen from Burdett with the above daily maximum and monthly maximum volume limitations.

TABLE C—22. ESTIMATED OXYGEN PLANT CAPITAL AND OPERATING COSTS

Oxygen uses	O_2 required metric tons/day (tons/day)	Source and type of plant	Oxygen purity (percent)	O_2 plant size metric tons/ day (tons/day)	Capital costs*	Operating costs $/ton*+#
Bleach plant	4.55—6.37 (5—7)	Purchased LOX	99.4	———	40,000	31.43
Bleach plant	4.55—6.37 (5—7)	Cryogenic O_2 plant	99.5	45.4 (50)	736,000	7.30
Waste treatment	26.4—40 (29—44)	Cryogenic O_2 plant	99.5			
Bleach plant	4.55—6.37 (5—7)	Cryogenic O_2 plant	99.5	58.6 (60)	795,000	6.83
Waste treatment	26.4—40 (29—44)	Cryogenic O_2 plant	99.5			
Bleach plant	4.55—6.37 (5—7)	Purchased LOX	99.5	———	———	39.45
Waste Treatment	26.4—40 (29—44)	PSA oxygen plant	95	Two at 27.2 (30)	1,350,000	9.14
Black liquor oxidation	33.7—38.2	PSA oxygen plant	95			
Bleach plant	4.55—6.37 (5—7)	Cryogenic O_2 plant	99.5	69.9 (77.1)	980,000	6.52
Waste treatment	25.5—40 (28—44)	Cryogenic O_2 plant	99.5			
Black liquor oxidation	33.7—38.2	Cryogenic O_2 plant	99.5			
Bleach plant	4.55—6.37 (5—7)	Cryogenic O_2 plant	99.5	82.3 (90.7)	1,095,000	6.47
Waste treatment	25.5—40 (28—44)	Cryogenic O_2 plant	99.5			
Black liquor oxidation	33.7—38.2 (37—42)	Cryogenic O_2 plant	99.5			
———	———	Pressure swing adsorption (PSA)	95	Two at 31.8 (35)	1,400,000	12.40
———	———	Cryogenic	99.5	57.6 (63.5)	1,500,000	10.01

*Costs are based on 1972 dollars.

+Based on 7 mills/kWhr power costs and 15-year straight line depreciation.

#Cost of Capital and LOX sales are not included in the operating costs.

£Cost of 34.1 cubic meter LOX storage tank and ambient vaporizer.

Process Description*

Chesapeake's cryogenic oxygen plant is best described by dividing the total system into four distinct process areas. These areas are: (a) air compresion, (b) air chilling, (c) air drying, and (d) air separation Figure C—23 is a basic flow diagram of the total process.

Air Compression—

Approximately 16,510 standard m^3/hour (583,000 scfh) of air enters the plant through a two-stage intake filter/silencer and flows to the first stage of a four-stage centrifugal compressor. A non-contact intercooler utilizing cooling water from a cross-flow cooling tower follows each of the first three compressor stages.

Compressed air at 1102 kPa gage (160 psig) is discharged from the fourth compressor stage to a non-contact after-cooler. The after-cooler utilizes water from the cross-flow cooling tower to remove the fourth stage heat of compression. Air temperature leaving the after-cooler is approximately 37.8° C (100° F). A mist extractor follows the after-cooler to remove entrained water droplets.

The four-stage centrifugal air compressor is driven by a 2500 horsepower, 1800 rpm direct-coupled, a-c, squirrel cage induction motor. The back end of the motor shaft is extended and directly connected to the main oil pump, which supplies lubrication to the air compressor gear box and motor bearings. An auxiliary oil pump with electric drive and electrical interlocks automatically operates during startup of the air compressor and at low oil pressure. An indirect oil cooler utilizing cooling tower water is provided to maintain an operating oil temperature of 43.3 to 48.9° C (110 to 120° F).

Controls for the compressor consist of a constant pressure control system to maintain a constant fourth-stage discharge pressure, a surge control system with atmospheric vent and silencer, and a motor overload protection system. The compressor automatically shuts down in the event of: (a) low oil pressure, (b) high oil temperature, (c) high air temperature to the fourth compressor stage, (d) execessive vibration, and (e) a power interruption exceeding 20 cycles duration. An alarm system is provided to indicate any of the above shutdown conditions and to indicate: (a) high vibration, (b) auxiliary lube oil pump operating, (c) high motor bearing temperature, and (d) high condensate level in any of the three intercoolers.

Air Chilling—

Compressed air from the mist extractor is further cooled to a temperature of approximately 10° C (50° F) in a non-contact heat exchanger called the air chiller. The cooling water required by the heat exchanger is chilled by direct contact with a portion of the waste nitrogen gas product. Dry waste nitrogen gas from the air separation system main heat exchanger enters the bottom of a packed column and progresses upward countercurrent to the falling cooling water. Humidified waste nitrogen gas exists the top of the column and is vented to atmosphere. Cold water is pumped from a reservoir at the bottom of the column to the air chiller. The heated cooling water from the air chiller is fed into the top of the column. Cooling water flow to the air chiller is controlled by the temperature of the return water.

After the process air is chilled, the compressed air enters a three-stage separator. This separator is designed to removed entrained water prior to the air dryers.

Air Drying—

In order to prevent freezing and subsequent plugging of the air separation system, all carbon dioxide and water must be removed from the chilled air. Air drying is achieved via adsorption on molecular sieve in a two-bed air dryer system. The operating cycle consists of one bed in operation

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*Process covered by U.S. Patent 3,070,966, held by Superior Air Products Co.

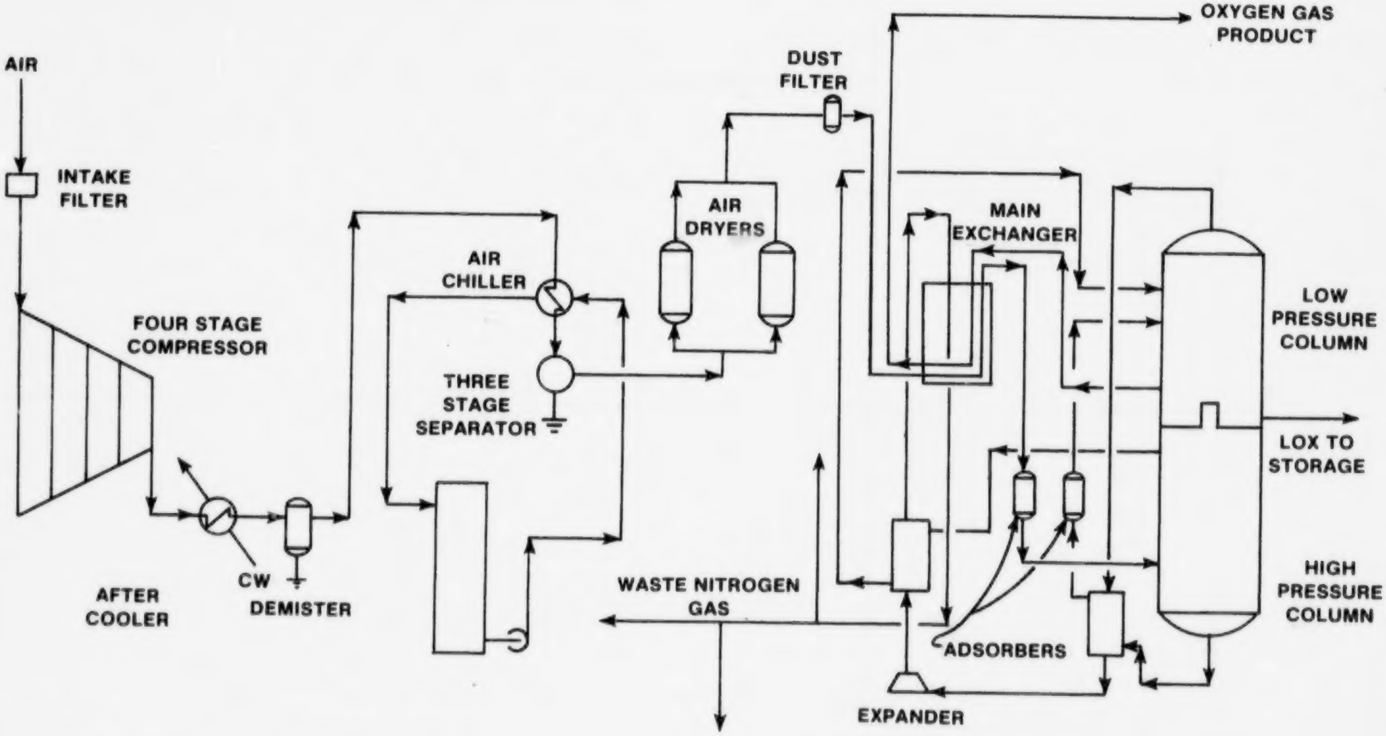

Figure C—23. Oxygen plant flow diagram.

while the other bed is being regenerated by waste nitrogen gas. Duration of both the operating and regeneration cycle is eight hours. Regeneration consists of: (a) depressurization, (b) heating, (c) cooling, and (d) repressurization.

Waste nitrogen gas for regeneration is heated to approximately 288° C (550° F) by a 140 kW electric heater and fed through the dryer bed. The duration of the heating cycle is maintained at three hours to ensure that carbon dioxide and water are expelled from the molecular sieve. At the conclusion of the heating cycle, the heater is automatically shut off and cool waste nitrogen at approximately 10° C (50° F) is fed through the dryer bed. Duration of the cooling cycle is four hours. During regeneration a maximum waste nitrogen temperature of 232° C (450° F) is discharged from the bed.

A gas blower preceded by a non-contact gas cooler and mist extractor were initially installed to convey the regeneration gas from the dryers to an atmospheric vent. Presently, as a result of an operating problem, the regeneration gas from the dryer bed is directly vented to atmosphere.

Carbon dioxide and water free air from the dryers discharge into a dust filter. The filter removes suspended particulate matter from the air stream prior to air separation.

Air Separation—

After air drying, the carbon dioxide and water free process air passes through the main economizer or heat exchanger. The process air is cooled in this heat exchanger to approximately -153° C (-243° F) by heat interchange with the waste nitrogen gas and the gaseous oxygen product. Cold process air from the economizer then passes through a silica gel gas adsorber for the removal of hydrocarbons. Following the adsorber, the process air is fed into the bottom of the high pressure distillation column. Every three weeks the silica gel adsorber is regenerated with heated waste nitrogen gas.

In the high pressure column, the process air is liquified and separated by distillation into an oxygen-rich liquid stream at approximately 35 percent (vol.) oxygen and a nitrogen-rich liquid stream at approximately 5 percent (vol.) oxygen. Operating pressure within the column is maintained at 882 kPa gage (128 psig). The oxygen-rich liquid stream is discharged from the bottom of the high pressure column and flows through an economizer where heat is interchanged with the cold waste nitrogen gas from the top of the low pressure distillation column. The oxygen-rich liquid at approximately -187° C (-304° F) is then passed through additional silica gel adsorbers for removal of acetylene. Following the adsorbers, the oxygen-rich liquid is expanded into the center of the low pressure distilation column. Every three weeks the silica gel adsorbers are regenerated with heated waste nitrogen gas.

The nitrogen-rich liquid stream is discharged from the top of the high pressure column and is passed through an economizer where heat is interchanged with cold waste nitrogen gas from the turbo expander outlet. The cold nitrogen-rich liquid at approximately -187° C (-304° F) is then expanded into the top of the low pressure column. Operating pressure within the low pressure column is maintained at 234 kPa gage (34 psig).

The nitrogen reflux required to liquify and fractionate the incoming air in the high pressure column is produced by condensing the nitrogen-rich gas from the top of the high pressure column. The refrigeration required for the nitrogen condensation is provided by evaporation of liquid oxygen in the low pressure column. Nitrogen-rich gas from the top of the high pressure column is condensed in an economizer at the bottom of the low pressure column. The condensed nitrogen reflux is then returned to the top of the high pressure column.

A high purity liquid oxygen product, a high purity gaseous oxygen product and a nitrogen-rich waste gas are produced by distillation in the low pressure column. Liquid oxygen is discharged from the bottom of the column and is cooled in an economizer to approximately -180° C (-292° F) by waste nitrogen gas from the turbo expander. From the economizer, liquid oxygen flows to a 106 m^3 (28,000 gallon) liquid oxygen storage tank. Flow of liquid oxygen from the low pressure column is controlled by liquid oxygen level in the low pressure column.

Gaseous oxygen is discharged from the bottom section of the column and flows into the main economizer, where heat is interchanged with the incoming process air. A flow control system maintains the gaseous oxygen flow from the economizer to the oxygen distribution system. Temperature and pressure of the gaseous oxygen product from the economizer are respectively 7.2° C (45° F) and 207 kPa gage (30 psig).

The nitrogen-rich waste gas stream is discharged from the top of the low pressure column and flows into an economizer where heat is interchanged with the oxygen-rich liquid from the high pressure column. From the economizer, the waste nitrogen gas flows into a turbo expander, where the gas is expanded from a pressure of 220 kPa gage (32 psig) to a pressure 34.5 kPa gage (5 psig). this expansion provides the refrigeration for the entire process. Flow to the turbo expander is controlled by a pressure control system on the low pressure column. The power developed by the turbo expander is absorbed and rejected as heat in an integral oil-filled dynamometer. An oil cooler utilizing cooling water from the cross-flow cooling tower removes the rejected heat from the dynamometer oil reservoir.

From the turbo expander, the waste nitrogen gas flows through three economizers, where heat is interchanged with the liquid oxygen product, the rich nitrogen liquid and the incoming process air. Following the main exchanger or economizer, a portion of the waste nitrogen gas is used for air chilling, a portion is used for regeneration of the air dyers, a portion is periodically used for regeneration of the silica gel adsorbers, a portion is vented directly to atmosphere and a fractional quantity is used to pressurize the air separation system cold box. All air separation equipment, excluding the turbo expander, is located within a heavily insulated structure called the cold box. This structure is pressurized with the dry waste nitrogen gas to prevent air leakage into the insulated system.

All oxygen plant cooling water, except for the air chiller, is supplied by a conventional cross flow cooling tower. Table C—23 summarizes the cooling tower water requirements for the oxygen plant.

Plant Startup and Operating Problems

A consolidated control room is provided for the oxygen plant and the oxygen activated sludge wastewater treatment plant. Operating personnel consists of a full-time operator and a helper. The operator is responsible for operation of both the oxygen and wastewater treatment plants. During normal operation, the oxygen plant requires very little operator attention. Routine work consists of: (a) measuring the oxygen product gas purity once per shift, (b) measuring the liquid oxygen acetylene concentration once per shift, (c) measuring the zinc, chloride and residual chlorine concentrations in the chilling water and cooling tower water twice a week, (d) taking plant instrumentation readings every four hours, and (e) touring the plant facilities every two hours.

The helper is primarily responsible for operation of the sludge dewatering facilities and transporting the dewatered sludge to the landfill. He also assists the operator when thawing the oxygen plant. During the yearly defrost of the entire oxygen plant, an off duty operator is called in to assist. Supervision is provided for the combined oxygen and wastewater treatment plants during the day shift five days a week.

Continuous operation of the oxygen plant began on May 5, 1974. Very few operating problems were associated with the startup. The following is a summary of the major operating problems:

During startup, the waste nitrogen gas blower in the air dryer regeneration cycle was overpressurized and exploded. The blower normally operates only during the heating and cooling sequence in the dryer regeneration cycle. Operation of the blower results in a slight vacuum at the dryer regenerant discharge line. The blower discharges the waste nitrogen regenerant directly to atmosphere. Due to the low operating pressure in the heating and cooling sequence, the waste nitrogen blower is rated at only 34.5 kPa gage (5 psig). Apparently due to a control failure, the blower was exposed to full system pressure, which caused the explosion.

Chesapeake elected not to install a new gas blower because of operator safety considerations. As a result, the turbo expander back pressure is maintained at 34.5 kPa gage (5 psig) instead of the design back pressure of 15.8 kPa gage (2.3 psig). This increased pressure is necessary to achieve the required regeneration gas flow to the air dryers. A small percentage of the refrigeration capacity and plant production capacity is lost due to operating without the waste gas blower.

Another operating problem was plant noise during thawing and during venting of oxygen from the oxygen distribution system. This would not be a problem at a normal industrial plant site. However, Chesapeake's oxygen plant is located approximately 100 meters (328 feet) from residential housing. To correct this problem, two additional gas silencers were purchased and installed.

Corrosion problems occurred in the mist extractor which follows the after cooler and in the three stage separator. The carbon steel vanes in the mist extractor and the carbon steel filter supports in the three stage separator were replaced with stainless steel.

A rigid coupling between the air compressor dirve and lube oil pump was replaced by a flexible coupling. The rigid connection resulted in compressor motor bearing failures.

TABLE C—23. COOLING WATER REQUIREMENTS (70)

Cooling water user	User requirement L/s	(gpm)
First stage intercooler	16.5	(262)
Second stage intercooler	16.5	(262)
Third stage intercooler	16.5	(262)
Fourth stage aftercooler	9.5	(150)
Compressor oil cooler	2.5	(40)
Waste nitrogen cooler (no longer used)	5.7	(90)
Expander dynamometer (71)	3.8	(67)
Expander oil cooler (71)	0.45	(7.2)
Total	71.5	(1,140.2)

The major operating problem occurred in the three stage separator and air dryers. Presure drop across the air dryers under normal operation is approximately 6.89 kPa gage (1 psig). On several occasions, the pressure drop in one or both of the air dryers has increased. This increased pressure drop affects operation by decreasing the plant air flow capacity. This in turn decreases the oxygen production capacity. In addition, regeneration of the air dryers is restricted by a reduction in the waste nitrogen gas flow.

Inspection of the dryers revealed that a portion of the molecular sieve had turned to a powder. This powder was partially plugging the inlet and discharge screens and thus causing the increased pressure drop. To eliminate the powder from the dryers, the entire molecular sieve contents of the bed were removed. The inlet and discharge screens were cleaned and the existing molecular sieve from the dryer bed was screened. The sieve material retained by the screen was returned to the bed and new molecular sieve was added to bring the bed to the design level.

On January 29, 1977, this problem recurred with severe consequences. Immediately following a dryer change, the air temperature entering the main economizer (from the air dryer) was excessive. The temperature on the strip chart recorder was off-scale with an inlet temperature in excess of 93° C (200° F). Pressure drop across the new dryer in service was approximately 138 kPa gage (20 psig). Plant gaseous production decreased and liquid oxygen production totally ceased.

The plant was shut down to determine the cause and extent of damage. Inspection of the dust filter (following the air dryers) revealed that the filter was destroyed by fire. Without the dust filter, all carryover from the air dryers, including debris from the filter fire, discharged directly into the main economizer. This resulted in partial plugging of the main economizer, causing inefficient heat transfer. Under normal operation, the temperature difference between the incoming air and gaseous oxygen product is approximately 2.8° C (5° F). After the filter fire, the main economizer warm end temperature difference was 13.9° C (25° F).

Superior Air Products assisted in cleaning the main economizer(72). Cleaning was accomplished by blanking off one of the two air inlets into the economizer and installing a pressure rupture disc on the other inlet. The main exchanger was then pressurized from the inside using the thawing inlet line to the gas adsorber. After the disc ruptured, a new disc was installed on the other inlet. The first rupture disc on each of the two inlets ruptured at 310 kPa gage (45 psig). Three more rupture discs were installed on each of the two inlets and ruptured at approximately 689 kPa gage (100 psig).

Inspection of the dryer in service revealed a light crust on top of the bed and an unlevel bed. The other dryer was also inspected, which revealed nothing abnormal about the bed. Insufficient time, due to a reduced inventory of liquid oxygen, was available to remove, clean and replace the molecular sieve in the damaged dryer.

The main heat exchanger was thawed overnight and plant production was commenced the following day. Warm end temperature difference across the main heat exchanger was 3.3° C (6° F), which indicated that the exchanger was clean.

Superior Air Products indicated that liquid water from the three stage separator was the probable cause of the deteriorated molecular sieve(73). A program was initiated to periodically check the water level in the three stage separator and to check the moisture removal efficiency of the aftercooler and three stage separator. After many months of normal operation, the three stage separator was found one day with an excessive water level. This high level only lasted for a few minutes. Inspection of the drain line revealed pipe scale partially plugging a globe drain valve. The drain line was replaced with a larger diameter stainless steel line and the globe valve was replaced by a stainless steel ball valve. This corrected the water carry-over problem.

In order to protect the main heat exchanger, the dust filter media material was changed. The new media material is rated at a maximum operating temperature of 427° C (800° F). This is compared to the previous media material maximum temperature rating of 121° C (250° F). In addition, the plant operator now inspects the gas dryer approximately thirty minutes prior to a dryer change. This is to ensure that the regenerated dryer is adequately cooled prior to change over.

Another problem is that the total system pressure drop form the air compressor fourth stage to the high pressure column is approximately 68.9 kPa gage (10 psig) in excess of the design pressure drop. An undersized pipe line is the primary cause of the excessive pressure drop. However, the plant meets production requirements and nothing to date has been done to correct this problem.

Finally, power interruption in excess of 20 cycles duration have frequently shut down the oxygen plant during the summer months. Thunderstorms within the immediate mill vicinity temporarily interrupt the purchase power distribution system and thus cause the shutdown. After several shutdowns, the oxygen plant must be thawed. A new power distribution system by Virginia Electric and Power Company has corrected this problem.

Plant Production

Gaseous oxygen production by Chesapeake's cryogenic plant is dependent on the total mill oxygen demand, which in turn is a function of the pulp mill production level. Table C—24 summarizes the 1976 oxygen plant production levels. This table is based on 13 production periods per year, with each period of 28 days' duration.

Average daily gaseous oxygen production is 1976 was 65.3 metric tons/day (72.0 tons/day) or 80 percent of the plant gaseous design capacity of 81.6 metric tons/day (90.0 tons/day). In order to operate the plant at a reduced gaseous production level, the plant air flow must be turned down. The oxygen plant, however, can only be turned down to approximately seventy-five percent (75 percent) of design capacity without venting oxygen product. Below 75 percent of design capacity, oxygen must be vented to prevent compressor surging and/or weeping of oxygen on the distillation trays(74).

Liquid oxygen production during 1976 averaged 4.48 metric tons/day (4.94 tons/day). This is approximately 49.4 percent of the plant design capacity. This low liquid oxygen production level reflects: (a) liquid oxygen transferred to the bleach plant LOX storage tank, (b) oxygen plant shutdowns, both scheduled and unscheduled, and (c) reduced liquid oxygen production capacity due to air dryer operating problems. During either scheduled or unscheduled oxygen plant shutdown, gaseous oxygen demand is met by vaporizing liquid oxygen from the 106 m^3 (28,000 gallon) liquid oxygen storage tank.

An acceptance test was conducted on the oxygen plant during the period from May 21 to May 25, 1974, to determine if the plant met the performance guarantee(75). This test was conducted without the waste nitrogen gas blower. Initially, the oxygen plant was accepted by Chesapeake Corporation based on the results of the acceptance test. However, the test was later invalidated due to the discovery that instrumentation installed to measure the oxygen production rate was faulty and gave results grossly in error(76). Another test has not been run because Chesapeake will not run the plant with the waste nitrogen gas blower and Superior Air Products will not accept the test results unless the blower is in operation. In general, the plant will produce the design capacity of 81.6 metric tons/day (90 tons/day) of gaseous oxygen and 9.07 metric tons/day (10 tons/day) of liquid oxygen at a purity of 99.5 percent except during extremely hot and humid days.

A test was conducted in October, 1976, to determine the plant production capacity at lower oxygen purities. This test was conducted during a period of high pressure drop across the three stage separator and subsequent low production level(77). Prior to the test, the plant production was 74.4 metric tons/day (82 tons/day) of gaseous and 9.1 metric tons/day (10 tons/day) of liquid oxygen with a product purity of 99.5 percent. At a product purity of 97.55 percent, the plant produced 85.3 metric tons/day (94 tons/day) of gaseous oxygen and 7.7 metric tons/day (8.5 tons/day) of liquid oxygen. The plant air flow was further increased to a production level of 88.0 metric tons/day (97 tons/day) of gaseous oxygen and 7.3 metric tons/day (8.0 tons/day) of liquid oxygen with a product purity of 95.0 percent.

Oxygen Plant Costs

Total installed cost of the 90.7 metric tons/day (100 tons/day) cryogenic oxygen plant was $1,260,000(78). Financing of this facility was by Pollution Control Bonds with an average yield to maturity of 5.9 percent.

Table C—25 summarizes the 1976 and 1977 operating costs for the oxygen plant. This table does not include the cost of capital in the operating expenses. Also, the revenue from liquid oxygen sales is not included.

TABLE C—24. OXYGEN PLANT PRODUCTION—1976

Production period	Total oxygen plant production (metric tons)	Liquid oxygen sold (metric tons)	Liquid oxygen purchased (metric tons)
1st	1706.15	18.58	5.35
2nd	2085.74	156.85	0
3rd	1916.79	17.36	0
4th	1985.31	110.61	0
5th	2053.29	210.76	0
6th	2005.29	155.81	0
7th	1514.56	96.45	0
8th	1878.00	93.62	0
9th	1670.16	41.70	0
10th	1825.90	202.58	0
11th	1772.22	156.40	0
12th	1912.14	234.17	0
13th	1450.14	136.16	52.07
Period average	1828.90	125.47	4.42

It should be noted that oxygen production, both liquid and gaseous, average 71.7 metric tons/day (79 tons/day) in 1976 and 68 metric tons/day (75 tons/day) in 1977. At these production levels the cost of oxygen was $22.44/metric ton in 1976 and $28.66/metric ton in 1977. A portion of the increase in oxygen cost per metric ton of production in 1977 is due to an increase in the power, material, and labor costs. In addition, the reduced production level in 1977 results in a higher oxygen cost per metric ton.

WASTEWATER TREATMENT SYSTEM

As indicated in Section 5 of this report, Chesapeake sought information on at least three different wastewater treatment systems. They included (a) activated sludge with air, (b) activated sludge with high-purity oxygen, and (c) aerated lagoons with secondary clarifiers. All three systems had sludge dewatering facilities. The system installed was chosen based on process and economic data described below.

System Selection

Chesapeake faced certain constraints in choosing the final waste water treatment system. Lack of acreage on the plant site placed the aerated lagoons about one mile from the plant boundaries but still within town limits; also residential areas were near the proposed site. The need to pump up to 61,777 m^3/day (16.3 mgd) to and from the lagoon system, concern for blowing foam and presence of odor near the residential areas and the anticipated difficulty in receiving citizen and town approval weighed heavily against this system.

TABLE C—25. OXYGEN PLANT OPERATING COSTS

Account	1976 Expenses	1977 Expenses
Power*	$344,590	$425,956
Supplies	19,652	35,080
Purchased oxygen	1,917	9,941
Repairs—labor and material	29,403	22,533
Production labor+	55,058	77,000
Depreciation	119,587	119,587
Total cost/year	$570,207	$690,097
Total production#	25,407 metric tons	24,077 metric tons
Oxygen cost/metric ton	$22.44 /metric ton	$28.66 /metric ton

*Based on 1976 power costs of $21.9/1000 kWh and 1977 power costs of $29.2/1000 kWh.

+Based on 35 percent of total production labor for oxygen plant and wastewater treatment plant.

#Includes gaseous and liquid oxygen production.

TYPING GUIDE SHEET

Acreage was available on the plant site for activated sludge using either air or high-purity oxygen. However, the acreage boundary lay within 45.8 m (150 feet) of several residences. Fearing foam and odor problems at times from an open biological system, such as activated sludge treatment with air, Chesapeake decided to investigate the vendors appealing claims for AST with high-purity oxygen. These claims, compared with air AST were (a) lower capital and operating costs, (b) less land space required, (c) no odor or foam problems from the closed bioreactor, (d) sludge that settled faster to a higher underflow concentration, (e) less waste sludge to dispose, and (f) much higher oxygen utilization efficiencies. There being no process data for high-purity oxygen AST on pulp and paper mill wastes, Chesapeake elected to carry out a pilot plant program to acquire such data and develop a process design which could be compared with air AST and aerated lagoon systems.

UNOX System Pilot Plant Study—

The steps taken by Chesapeake to purchase and operate a 0.114 m^3/min (30 gpm), 4-stage pilot unit to test oxygen AST are described briefly in Section 5. The actual pilot study was completed on May, 1972 prior to the beginning of the grant period. However, a general description of the pilot plant facilities and the pilot plant study program follows. Details of the pilot studies have been reported in the literature.(79)(80)

The trailer mounted pilot plant started up in September 1971 and operated in different modes for 8 months until early May 1972. The unit received four streams pumped from different sewer systems in the mill. These were waste streams from the pulp mill, paper mill, and the recovery room; the fourth stream was condensate from the evaporators. A schematic flow diagram of the pilot unit is shown in Figure C—24.

The general pilot plant program was organized and conducted as indicated in Table C—26, which shows the different loadings and study periods.

During the pilot studies the raw waste streams were blended and pumped to the primary clarifier designed for 1.1 m^3/m^2-hr (650 gpd/ft^2) and 4 hours' detention time at 0.133 m^3/min (35 gpm); the primary clarifier had no rake mechanism. Piping was provided to bypass the primary clarifier for the Phase V study.

The heat exchanger cooled the primary supernatant to 35° C (95° F) which then flowed to a 0.57 m^3 (150 gallon) tank where defoamer and nutrients (NH_3 and H_3PO_4) were added. The mixture flowed then to the biological reactor consisting of 4-stages of 1.52 m^3 (400 gal) each and hydraulic

TABLE C—26. GENERAL PILOT PLANT PROGRAM

Phase	Study period	Loading*
I	October 3—November 7, 1971	0.75
II	November 14—December 18, 1971	0.54
III	January 9—February 5, 1972	1.30
IV	February—March 1972	Sludge dewatering
V	April 19—May 11, 1972	Operation with primary clarifier bypassed

*Loading is kilograms (pounds) of BOD applied per day per kilogram (pound) of volatile suspended solids (MLVSS).

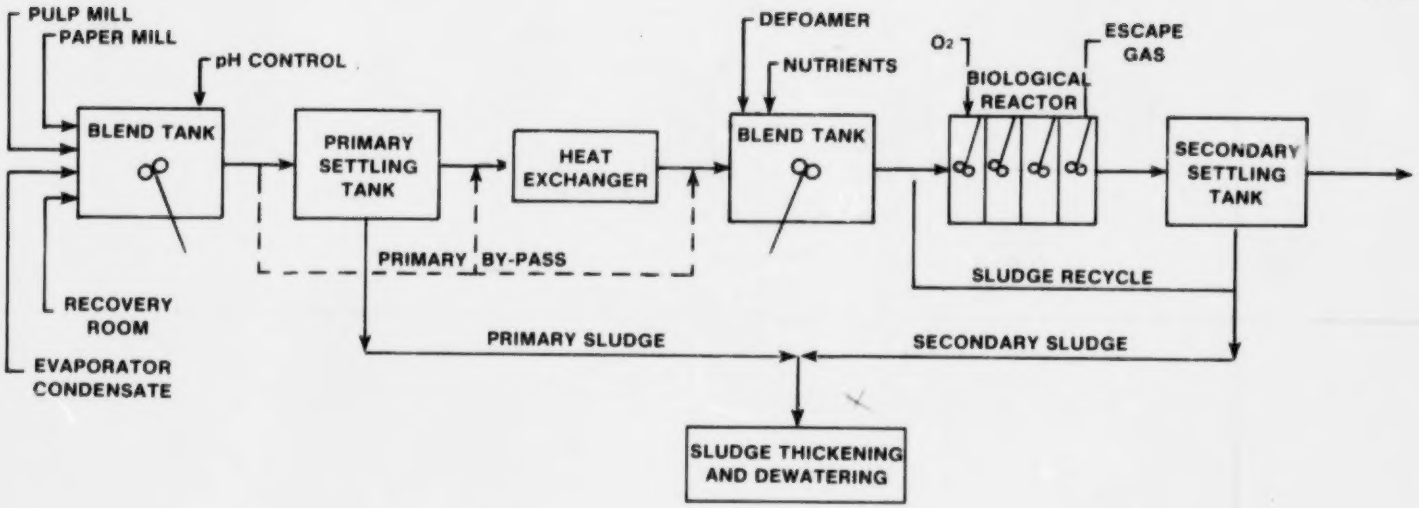

Figure C—24. Schematic flow diagram of pilot treatment system.

detention times of 1.3 to 2.3 hours. MLVSS ranged from 4,000 to 7,000 mg/L. High-purity oxygen entered the first stage and passed cocurrently in series to each of the other three stages with the waste blend. Effluent flowed to a secondary clarifier which was operated at overflow rates of 0.85 to 1.53 m^3/m^2-hr (500 to 900 gpd/ft^2). Secondary sludge was recycled to the first stage of the bioreactor and excess sludge was wasted when necessary to maintain the sludge blanket in the secondary at the level desired. Pilot sludge dewatering studies were carried out on a 0.91 m^2 (10 ft^2) vacuum coil filter.

Chesapeake operated the pilot system seven days per week and collected and analyzed all samples. The samples were composited, stored at 4° C (39.2° F) before laboratory analysis, and were analyzed seven days per week.

Results of the pilot plant studies are summarized in exerpts from descriptions of the pilot plant studies in the literature. The initial report(79) of pilot plant results stated that the pilot plant UNOX program at Chesapeake had definitely determined that it is feasible to effectively treat pulp and papermill wastes using high purity oxygen in the activated sludge process. The UNOX system pilot plant was consistently operated at high MLSS levels (5000 to 7000 mg/L), at high DO concentrations, ($>$ 5 mg/L), high overall utilization of oxygen feed gas ($>$ 90 percent), and high secondary system BOD_5 removal of greater than 90 percent. Settling and thickening properties of the secondary sludge produced were good and the amount of secondary sludge produced was low in compariosn to common literature values. Dewatering of the secondary sludge on a vacuum filter at low primary to secondary ratios and at low feed consistencies was successful without chemical conditioning. There was not a decrease in sludge setteability as the food to micro-organism ratio increased to over 1.3 and the organic loadings reached values of 96 g/L (600 lb BOD_5/1000 ft^3) of tank volume. However, turbidity increased somewhat during the peak organic loadings. This turbidity increase is not peculiar to the UNOX system, but rather to biological waste treatment.

The increased buffering capacity of the UNOX system because of higher dissolved carbon dioxide levels did not cause problems in treating unbleached kraft pulp and paper mill wastes. However, lower alkalinity wastes such as those from fully bleached operations may require caustic to control pH in the proper range for good biological treatment.

A later report gave a fuller description of the sludge settling and dewatering characteristics drew similar conclusions.(80)

This report indicated that the high purity oxygen system was operated at organic loadings ranging from 0.54 to 1.3 lb of BOD removed per day per pound of MLVSS. The treatment system consistently provided good removals and produced a biological sludge with excellent settling and thickening characteristics. This high rate system was able to maintain operational stability over a wide range of loading conditions. Short-term stock loads of BOD_5 and suspended solids had little effect on the normal operation of the plant.

Settling and thickening characteristics of the primary and secondary sludges were also evaluated. Laboratory studies on the primary underflow indicated that the sludge could be thickened in the clarifier to 3 percent. Pilot plant and laboratory results demonstrated that the secondary sludge could be thickened in the secondary clarifier to 2.5 percent. Throughout this study, the biological sludge developed in the high purity oxygen system exhibited excellent settling and thickening characteristics.

Vacuum filter studies demonstrated that both the total suspended solids level and primary solids level in the feed have a significant effect on loading. Increasing the concentration of solids in the feed resulted in an increase in the allowable loading up to a limiting solids level. Above 3 percent solids, no appreciable change in loading was observed. Primary-to-secondary sludge ratios of 2:1 operated more efficiently than the 1:1 ratio as demonstrated by increased loading and cake solids. Bark addition examined at the 1:1 ratio resulted in slightly higher loadings than observed without bark addition. Sludge dewatering on the pilot filter indicated that higher cake solids could be obtained at the 2:1 ratio. Cake solids in the range of 18 percent were observed for 2:1 mixture, while a range of 12 percent of 16 percent was observed for the 1:1 mixture.

As a result of these pilot studies, it was concluded that oxygen AST was a viable process for treating kraft mill wastewaters and that the process design requirements of this and alternate conventional air systems should be established. In order to develop appropriate process design criteria, the quantity and nature of the expected influent to an external WWTP also needed to be defined. The work to define influent characteristics proceeded concurrently with the pilot studies.

A series of memoranda(81) in early 1972 examined mill parameters and projected the expected BOD_5 and TSS loads, the total flow, the wastewater temperature, the alkalinity, and pH for the WWTP influent. These memoranda, consisting of eighty-five pages of text, data, computations and graphs resulted in the projections shown in Table C—27.

Hydroscience Process Design Development—

Data from the pilot plant study and the projections listed in Table C—27 were used by Hydroscience to develop process design for the high purity oxygen AST. The Linde Division of Union Carbide also used the data to develop detailed design criteria for the bioreactors. The Hydroscience report(82) providing process design from the pilot plant data was submitted to Chesapeake in July 1972. This very comprehensive report was not published per se although substantive portions were included in the published literature.(79)(80)(83) The system recommended from the pilot study is shown schematically in Figure C—25. Recommended process design information is summarized in Table C—28.

TABLE C—27. EXPECTED WASTE TREATMENT PLANT LOADS

Type of load	Definition	Projected level	
BOD_5, kg/hr (lb/day)	Average		
BOD_5, kg/hr (lb/day)	Average	852	(45,000)
	Average peak	1,080	(57,000)
	Peak	1,326	(70,000)
Flow, m^3/day (mgd)	Average	61,780	(16.3)
Suspended solids kg/hr (lb/day)	Average (50 percentile)	484	(25,550)
	90 percentile	960	(50,710)
	Peak	1,407	(74,310)
Methyl orange alkalinity mg/L	Minimum	100	
	Average	428	
pH	Minimun	6.0	
	Average	8.6	
Temperature, °C (°F)	Average	62.8	(145)
	90 percentile		
	Peak	70.6	(159)

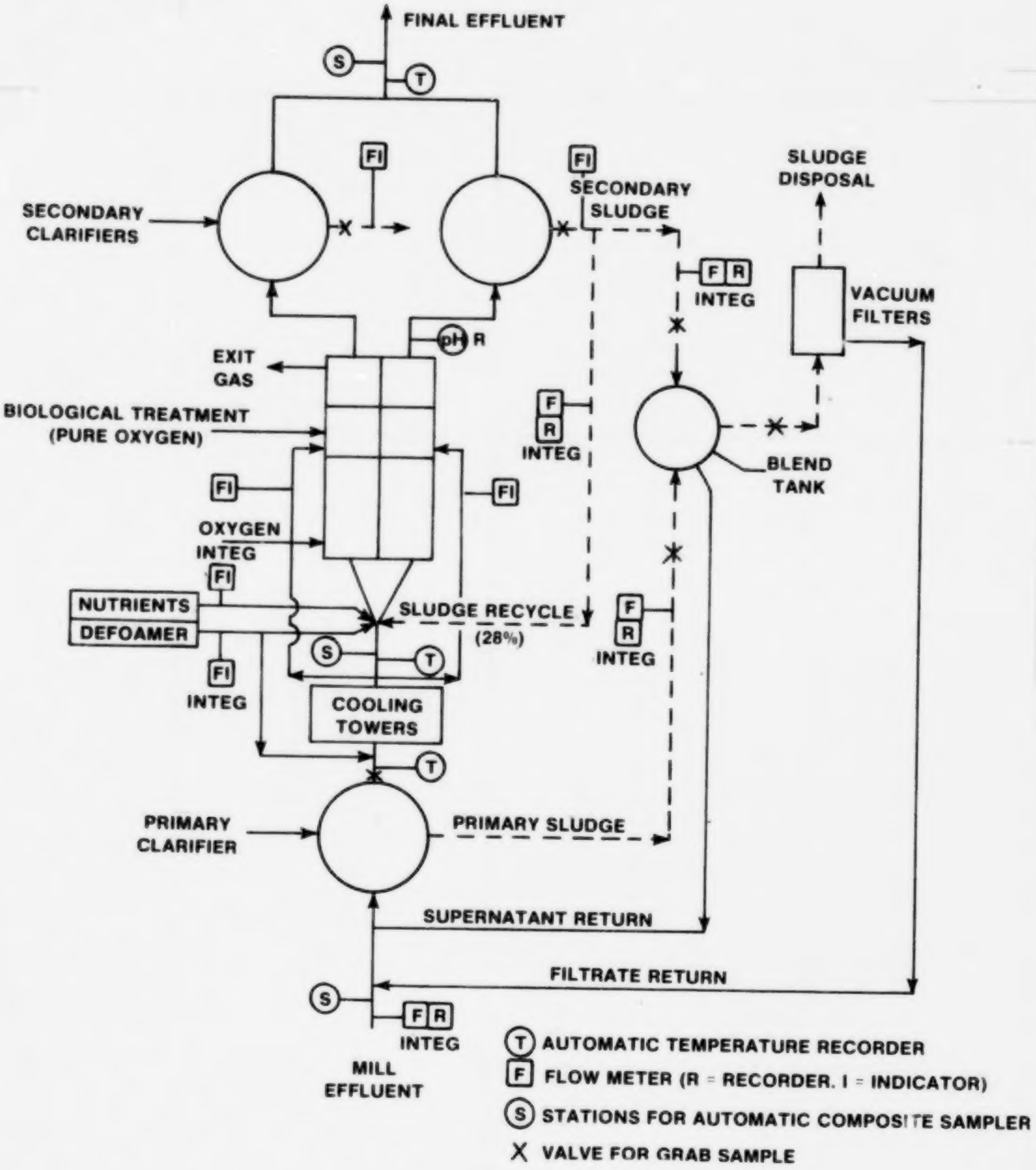

Figure C—25. Schematic flow diagram of recommended treatment system.

TABLE C—28. LIST OF PROCESS DESIGN INFORMATION

Item	Description
Waste load definition:	
Flow	61,684 m^3/day (16.3 mgd)
Average BOD load	20,412 kg/day (45,000 lb/day)
Peak BOD load	31,500 kg/day (70,000 lb/day)
Average solids load	11,475 kg/day (25,500 lb/day)
90 percent solids load	22,820 kg/day (50,710 lb/day)
Temperature (after precooling the mill effluent)	35°C (95°F)
Primary clarifier:	
Design basis (suspended solids removal and no BOD removal)	55 percent
Overflow rate	2.31 m^3/m^2/hr (1,360 gpd/ft^2)
Detention time	70 min
Design of primary clarifier	Controlled by clarification
Biological treatment system (pure oxygen):	
Organic load	0.68 kg/day BOD_R day/lb MLVSS (0.68 lb BOD/day/lb MLVSS)
MLVSS	4,900 mg/L
Sat	11.35 g/L-hr
Oxygen requirement (peak BOD load)	36,315 kg/day (80,700 lb/day)
Biological solids produced	2,745 kg/day (6,100 lb/day)
Sludge recycle rate	28 percent (average) 56 percent (peak)
Total secondary solids for disposal (average conditions)	4,860 kg/day (10,800 lb/day)
Total solids for disposal (average conditions)	11,160 kg/day (24,800 lb/day)
Total BOD removed	92 percent
Ammonia requirement	
Average conditions	716 kg/day (1,590 lb/day)
Peak conditions	1,859 kg/day (4,130 lb/day)
Phosphoric acid requirement	
Average conditions	373 kg/day (828 lb/day)
Peak conditions	968 kg/day (2,150 lb/day)
Defoamer requirement	
Minimum	284 kg/day (630 lb/day) or 0.33 m^3/day (86 gal/per day)
Peak	945 kg/day (2,100 lb/day) or 1.09 m^3/day (288 gal/day)

Secondary clarifier:	
Overflow rate	24.5 m^3/m^2/day (600 gpd/ft^2)
Clarification surface area	2,527 m^2 (27,200 ft^2)
Underflow concentration	2.5 percent
Allowable flux	166 kg/m^2/day (34 lb/day/ft^2)
Thickener surface area	2,620 m^2 (28,200 ft^2)
Blend tank:	
Underflow concentration	3.0 percent
Allowable flux	166 kg/m^2/day (34 lb/day/ft^2)
Vacuum filter:	
System operation	8 hr/day, 5 days/week
Solids load	15,615 kg/day (34,700 lb/day)
Loading	44 kg/m^2/hr (9.0 lb/hr/ft^2)
Sludge disposal:	
Method	Landfill
Percent cake	16 percent
Solids load	11,160 kg/day (24,800 lb/day) (7 day/week, 8 hr/day operation)
Solids load	15,615 kg/day (34,700 lb/day) (9 day/week, 8 hr/day operation)
Cooling Tower:	
Temperature in	
Average	60.6°C (141°F)
Peak	68.3°C (155°F)
Temperature out	
Design	35°C (95°F)

Hydroscience also used the data developed during the pilot study and their experience in pulp and paper WWTP design to develop process designs for a conventional air AST system and an aerated lagoon system. The sizing and economics of these three systems are compared in the following section.

Cost Estimates of Alternate Treatment Systems—

The comprehensive Hydroscience report(82) referred to previously discussed preliminary cost estimates of treatment alternatives. This report stated that preliminary sizing of three treatment systems was developed and cost estimates were made for each system to assist Chesapeake Corporation in the selection of the wastewater treatment process. The three biological treatment systems were:

I—Pure Oxygen Treatment
II—Conventional Activated Sludge
III—Aerated Lagoon

Tables C—29, C—30, and C—31, respectively, summarize the preliminary sizing of the treatment units in each system. Evaluation of process design criteria was sufficiently developed at this time period (May 11, 1972), to make valid cost comparisons of the alternatives. However, it will be noted that modifications have been made in the final recommended unit sizing. All are relatively minor, with the following exceptions:

(1) Subsequent settling studies demonstrated primary clarifier settling rates could be increased from the preliminary 1.02 m^3/m^2/hr (600 gpd/ft^2) to 2.31 m^3/m^2/hr (1,360 gpd/ft^2). Therefore, a reduction in primary tankage to one 38-meter (125-foot) diameter unit has been made.

(2) Subsequent coliform data demonstrated chlorination was not required. These preliminary cost estimates assumed chlorination for the three alternatives. It therefore, affected each cost estimate equally.

TABLE C—29. PRELIMINARY SIZING OF UNIT OPERATIONS PURE OXYGEN SYSTEM, May 11, 1972

Item	Description
Waste characteristics:	
Flow	61,777 m^3/day (16.3 mgd)
Average BOD_5	20.250 kg/day (45,000 lb/day)
Peak BOD_5	31,500 kg/day (70,000 lb/day)
Average S.S.	11,475 kg/day (25,500 lb/day)
Peak S.S.	22,820 kg/day (50,710 lb/day)
Primary clarifiers:	
Number of tanks	3
Diameter	33.6 m (110 ft)
Depth	3.1 m (10 ft)
S.S. removal	7,808 kg/day (17,350 lb/day)
Aeration tank:	
Volume	7,049 m^3 (1.86 mg)
Depth	3.81 m (12.5 ft)
Compartments	3 completely enclosed
Surface	8 aerators
Oxygen requirements	35,505 kg/day (78,900 lb/day)
NH_3-H requirements	590 kg/day (1,310 lb/day)
H_3PO_4 requirements	362 kg/day (805 lb/day)
Secondary clarifiers:	
Number of tanks	3
Diameter	33.6 m (110 ft)
Depth	20 percent (12,355 m^3/day) (3.26 mgd)
S.S. wasted	3,429 kg/day (7,620 lb/day)

Thickener:	
Gravity	176.5 m^2 (1,900 ft^2)
Diameter	15.3 m (50 ft)
Depth	3.1 m (10 ft)
Vacuum filter:	
Two coil units	23.2 m^2 each (250 ft^2)
Wet sludge for disposal:	
Dry solids	11,237 kg/day (24,970 lb/day)
Wet weight (16 percent solids)	70,200 kg/day (156,000 lb/day)
Volumetric rate	70.8 m^3/day (18,700 gal/day)
Chlorination:	
Volume	644.3 m^3 (170,000 gal)

TABLE C—30. PRELIMINARY SIZING OF UNIT OPERATIONS AIR ACTIVATED SLUDGE SYSTEM, MAY 11, 1972

Item	Description
Waste characteristics:	
Flow	61,777 m^3/day (16.3 mgd)
Average BOD	20,250 kg/day (45,000 lb/day)
Peak BOD	31,500 kg/day (70,000 lb/day)
Average S.S.	11,475 kg/day (25,500 lb/day)
Peak S.S.	22,820 kg/day (50,710 lb/day)
Primary clarifiers:	
Number of tanks	3
Diameter	33.6 m (110 ft)
Depth	3.1 m (10 ft)
S.S. removal	7,808 kg/day (17,350 lb/day)
Aeration tank:	
Volume	10,385 m^3 (2.74 mg)
Depth	4.6 m (15 ft)
Oxygen requirements	31,707 kg/day (70,460 lb/day)
Power	1,537 kW 92,060 hp)
Turbine aerators:	783 kW (1050 hp to turbine) 783 kW (1,050 hp to blowers)
NH_3-N requirements	986 kg/day (2,190 lb/day)
H_3PO_4 requirements	605 kg/day (1,345 lb/day)

Secondary clarifiers:	
Number of tanks	3
Diameter	33.6 m (110 ft)
Depth	3.1 m (10 ft)
Recycle rate (33-1/3 percent)	20,580 m³/day (5.43 mgd)
Gravity thickener:	
Diameter	18.3 m (60 ft)
Depth	2.44 m (8 ft)
Vacuum filter:	
Two coil filters	.41.8 m² (450 ft²)
Wet solids for disposal:	
Dry solids	10,530 kg/day (23,400 lb/day)
Wet weight (16 percent cake)	65,700 kg/day (146,000 lb/day)
Volumetric rate	66.7 m³ (17,600 gpd)
Chlorination:	
Volume	644.3 m³ (170,000 gal)

TABLE C—31. PRELIMINARY SIZING OF UNIT OPERATIONS AERATED LAGOON SYSTEM, MAY 11, 1972

Item	Description
Waste characteristics:	
Flow	61,777 m³/day (16.3 mgd)
Average BOD	20,250 kg/day (45,000 lb/day)
Peak BOD	31,500 kg/day (70,000 lb/day)
Average S.S.	11,475 kg/day (25,500 lb/day)
Peak S.S.	22,820 kg/day (50,710 lb/day)
Primary clarifiers:	
Not required	
Aerated lagoons:	
Volume	239,900 m³ (63.3 mg)
Surface area	69,675 m² (750,000 ft²)
Depth	5.4 m (12 ft)
Oxygen requirements	36,000 kg/day (80,000 lb/day)
Power	1,753 kW (2,350 hp)
Surface aerators	

NH_3-N requirements	765 kg/day 91,700 lb/day)
H_3PO_4 requirements	473 kg/day (1,050 lb/day)
Secondary clarifiers:	
Number of tanks	3
Diameter	33.6 m (110 ft)
Depth	3.1 m (10 ft)
Recycle rate	3,259 m^3/day (0.86 mgd)
Gravity thickener:	
Diameter	13.7 m (45 ft)
Depth	3.1 m (10 ft)
Vacuum filter:	
Two coil filters	27.9 m^2 (300 ft^2 each)
Wet sludge disposal:	
Dry solids	15,368 kg/day (34,150 lb/day)
Wet weight	72,000 kg/day (160,000 lb/day)
Volumetric weight	98.5 m^3/day (26,000 gal/day)
Chlorination:	
Volume	644.3 m^3 (170,000 gal)

Table C—32 presents the unit cost information used in developing the capital cost for each alternative. Capital costs did not include sewer segregation of the plant sewers, nor the outfall pipeline from the treatment plant to the discharge point in the Pamunkey River. In addition, no allowance was made for unusual foundation conditions.

TABLE C—32. BASIS FOR CAPITAL COSTS

Item	Unit Cost
Reinforced concrete	\$150/yd^3
Excavation	2/yd^3
Backfill	4/yd^3
Sheeting	4/yd^3
Handrailing	20/ft
Control building superstructure	15/ft^2
Site work	
Clearing and grubbing	600/acre
Roads	6/yd^2
Grating and seeding	0.80/yd^2
Fence	6/ft
Miscellaneous and contingencies	25 percent
Electrical	7.5 percent total capital cost
Instrumentation	15 percent mechanical equipment

Table C—33 presents the basic preliminary capital cost breakdown for the alternative treatment systems. The cost estimates were made for Hydroscience, Inc., by Nebolsine, Toth, and McPhee Associates. The cost of the aeration basin, aeration equipment, and instrumentation for the UNOX pure oxygen system was furnished by the vendor. The preliminary cost includes 25 percent for contingencies.

TABLE C—33. PRELIMINARY CAPITAL COST ESTIMATES ALTERNATE TREATMENT SYSTEMS (MAY 11, 1972)

Item	Alternates I (Oxygen)	II (conv. activ. sludge)	III (Aerated lagoon)
Primary clarifiers and sludge pumps	$774,000	$774,000	———
Aeration tanks	538,000	720,000	*641,000
Aeration equipment and instrumentation	———	405,000	640,000
(Unox)	+565,000	———	———
Installation (+25%) + (7-½%)	100,000	———	———
Secondary clarifiers and sludge pumps	808,900	808,900	808,900
thickener	93,600	109,100	85,100
Vacuum filter	322,500	488,200	322,500
Chlorination	143,000	143,000	143,000
Piping and valves (in-plant)	314,400	346,200	301,000
Site work	110,000	110,900	156,600
Control building			
3,200 ft²	236,000	———	236,000
7,200 ft²	———	491,400	———
Instrumentation	#108,000	154,700	154,700
Total	$4,115,100	$4,550,000	$3,489,000
Additional costs	Cooling	Cooling	Land and transmission

*Concrete lining of aeration basin.
+Cost does not include 25 percent contingency and 7-½ percent electrical.
#Unox instrumentation is included with aeration package.

The estimated operation and maintenance costs for each treatment system are shown on Table C—34. The basis for the estimated costs is presented. Sludge disposal costs are trucking costs to remove the sludge to landfill sites.

TABLE C—34. PRELIMINARY OPERATION AND MAINTENANCE COST ESTIMATES (MAY 11, 1972)

	Alternate Treatment Systems		
Item	**I (Oxygen)**	**II (Conv. activ. sludge)**	**III (Aerated lagoon)**
Maintenance	$13,900	$11,800	$14,000
Chemicals (1,2)	70,100	(3)124,000	(3)100,000
Power	38,200	88,000	96,000
Sludge disposal	75,000	101,000	69,000
Operators	55,000	55,000	55,000
Cooling tower and transmission(4)	47,000 (1,050 hp)	47,000 (1,050 hp)	27,000 (600 hp)
O_2 generating plant(5) (50 tons/day) 45 metric ton/day	79,000	———	———
Total	$378,000	$426,800	$361,300

(1) Neutralization costs not included.
(2) Cost of anti-foam not included.
(3) Includes $FeCl_3$, for sludge conditioning.
(4) Horsepower estimates by The Rust Engineering Company.
(5) Information from Chesapeake Corporation.

COST BASIS

1. Maintenance: 1 percent mechanical cost plus 0.1 percent civil construction costs.
2. Chemical costs:
   Ammonia $0.78/kg ($0.35/lb)
   H_3PO_4 20.67/100 kg ($9.31/100 lb)
   Chlorine $0.19/kg ($0.085/lb)
   $FeCl_3$ $8.89/100 kg ($4.00/100 lb)
3. Power costs: 7 mill/kW hr
4. Sludge disposal, landfill: $3.02/wet metric ton ($2.70/wet ton)

Table C—35 summarizes the capital costs and the operation and maintenance costs for the alternate systems. Cooling of the raw wastewaters before biological treatment is required for the pure oxygen and conventional activated sludge systems. No prior cooling is required aerated lagoon system since sufficient dissipation of heat can be obtained within the lagoon itself. The aerated lagoon system could not be located on the main mill grounds because of lack of sufficient land area. Greater transmission costs, therefore, are required to transport the wastewaters to the lagoon site. Cooling tower and transmission main costs were estimated by The Rust Engineering Company. Included in the pure oxygen system is the estimated cost of a 45 metric ton/day (50 ton/day) oxygen generating plant. The total preliminary capital costs range from $4.0 million for the aerated lagoon system to $5.3 million for the pure oxygen system. The latter includes $735,000 for the oxygen generating plant.

The yearly operation and maintenance costs ranged from $290,800 for the pure oxygen system, to $426,800 for activated sludge. The O&M cost shown for the pure oxygen system have been reduced by $88,000 to account for the yearly cost benefit that will be derived by the sale of excess liquid oxygen, and for furnishing the oxygen bleach plant.

Writing an account of Chesapeake's reasoning from these preliminary sizing and cost data is rendered somewhat difficult by other factors which influenced final decisions on treatment system

TABLE C—35. SUMMARY OF PRELIMINARY TREATMENT COSTS

	Alternate Treatment Systems		
Item	**I (Oxygen)**	**II (Conv. activ. sludge)**	**III (Aerated lagoon)**
Capital costs			
Treatment plants	$4,115,000	$4,551,000	$3,489,000
Cooling tower*	150,000	150,000	———
Transmission main*	313,000	313,000	572,000
O_2 generating plant (50 ton/day) (44.6 metric ton/day)†	735,000	———	———
Total	$5,313,000	$5,109,000	$4,061,000
Operation and maintenance (per year)	378,000	426,800	361,300
Credits † #	88,000	———	———
Total	$290,800	$426,800	$361,300

*Estimates by The Rust Engineering Company.
†Information from Chesapeake Corporation.
#Credits include supplying oxygen to oxygen bleach plant and sale of excess liquid oxygen.

selection. However, the reasoning recalled by several personnel involved closely approximates the following discussion.

Examining the capital cost data in Table C—35 it appears that aerated lagoons would be the lowest cost system to install. As mentioned earlier, however, this system has the disadvantage of (1) being located off the mill site in an area that would probably develop residentially and (2) the system would be completely open and subject to foam and odor formation. In spite of these clear disadvantages, the system was considered further because of its apparent lower cost as will be described later in this section.

The next least expensive system (Table C—35) was conventional air activated sludge, which was $204,000 cheaper than the pure oxygen system. However, giving credit to the oxygen system for oxygen supply to the bleach plant and income from liquid oxygen sales, a total of $88,000, made the annual operating cost of the conventional AST $136,000 higher than that of the oxygen system. Combining this factor with the fact that the AST system would require larger, open surface aeration volume and would, therefore, consume more of limited mill-site space located within less than 200 feet of existing residences, the conventional AST was eliminated from further consideration. The pure oxygen and aerated lagoon systems were compared again by Rust Engineering and Chesapeake as indicated in the remainder of this section and in the following section.

Tables C—36, C—37, and C—38 show detailed capital cost estimates for (1) the pure oxygen system with circular secondary clarifiers, (2) the pure oxygen system with rectangular secondary clarifiers (a variant included because of reported lower cost of rectangular clarifiers versus circular ones), and (3) the aerated lagoon system. The costs were assembled by Rust Engineering after lengthy discussions with Chesapeake and Hydroscience. Note, for example, when comparing estimated capital costs in Tables C—33 and C—36, that the totals for the pure oxygen system were almost the same but individual items differed substantially. The major differences in the individual items occurred apparently because Rust Engineering assembled direct construction costs (which were spread to each item by Hydroscience) into one lump sum. The same seems to be true when comparing Tables C—33 and C—38 for the aerated lagoon system.

The difference in capital cost between circular and rectangular secondary clarifiers can be seen in Tables C—36 and C—37. The rectangular clarifiers cost only slightly more than one third the cost of circular clarifiers. Therefore, this system was retained for further comparison.

Estimated operating costs in Table C—34 and C—39 differ by only $15,600 for the pure oxygen system (before taking credit for providing O_2 to the bleach plant and for the income from liquid O_2 sales) and by only $8,315 for the aerated lagoon system, both at the $350,000 to 360,000 level.

TABLE C—36. CAPITAL COSTS FOR UNOX SYSTEM (CIRCULAR SECONDARY CLARIFIERS)

Item	Cost
Lift pumps with piping	$251,049
Primary clarifier — 125'-0" diameter	63,865
Clarifier mechanism	80,000
Sludge pumps (primary)	11,970
Sludge pump pit	8,000

Cooling tower	
Cooling tower foundations	95,000
Cooling tower pumps	26,000
	27,950
Aeration chamber	325,600
Dissolution system	484,000
Secondary clarifiers (2 circular -135′ diameter)	172,238
Clarifier mechanism	194,000
Sludge pumps (secondary)	11,970
Sludge pump pit	12,000
Sludge thickener	80,880
Vacuum filter	167,000
Chlorination system	114,400
Electrical (2927 hp installed)	175,620
Instrumentation	197,097
Clearing and grubbing	4,000
Piping	364,805
Control building (3200 ft^2)	54,400
Direct construction cost	2,921,844
Indirect construction cost at 25 percent	730,973
Subtotal	$3,652,817
50-ton oxygen plant	$580,000
Installation (including contingency)	121,000
Storage tank	35,000
Subtotal	$736,000
Total cost	$4,388,817

TABLE C—37. CAPITAL COST FOR UNOX SYSTEM (RECTANGULAR SECONDARY CLARIFIERS)

Item	Cost
Lift pumps with piping	$251,049
Primary clarifier — 125 ft - 0 in. diameter	63,865
Clarifier mechanism	80,000
Sludge pumps (primary)	11,970
Sludge pump pit	8,000
Cooling tower	95,000
Cooling tower foundations	26,000
Cooling tower pumps	27,950
Aeration chamber	325,600
Dissolution system	484,000
Secondary clarifiers (2 rectangular 65 ft by 200 ft)	104,534
Clarifier mechanism	28,000
Sludge pumps (secondary)	11,970
Sludge pump pit	8,000
Sludge thickener	74,880
Vacuum filter	160,000
Chlorination system	114,400
Electrical (2927 hp installed)	175,620
Instrumentation at 7 percent equipment cost	166,500
Clearing and grubbing	1,000
Piping	332,805
Control building (3200 ft^2)	54,400
Direct construction cost	2,605,543
Indirect construction cost at 25 percent	651,904
Subtotal	$3,257,447
50-ton oxygen plant	1580,000
Installation (including contingency)	121,000

Storage tank	35,000
Subtotal	$736,000
Total Cost	$3,993,447

TABLE C—38. CAPITAL COST FOR AERATED LAGOON SYSTEM

Item	Cost
Lift pumps with piping	$545,843
Marsh crossing	4,600
Sludge pumps (primary)	11,970
Pump pit	8,000
Lagoons	231,200
Aerators	475,000
Aerator electrical	91,000
Lagoon lining	111,160
Secondary clarifier 2 at 135 ft	172,238
Mechanism	194,000
Sludge pumps (secondary)	11,970
Pump pit	8,000
Sludge thickener	68,080
Vacuum filter	194,000
Chlorination	114,400
Electrical (2762 hp)	165,720
Instrumentation	104,655
Clearing and grubbing	10,000
Piping	281,030
Control building	54,400
Direct construction cost	$2,857,266
Indirect construction cost (25 percent)	714,317
Total	$3,571,583

Chesapeake next proceeded to compare the two basic systems by a discounted cash flow computation. The method used is explained and the results obtained are presented in the following section.

TABLE C—39. OPERATING COSTS OF ALTERNATE WASTE TREATMENT SYSTEMS — MAY 12, 1972

Item	Aerated Lagoon	
	Oxygen	Cost
Maintenance	$13,900	$14,300
Chemicals		
Nutrients	45,500	59,500
Sludge conditioning	———	16,600
Chlorination	24,600	24,600
Power		
Aerators	25,350	82,375
Pumping (within treatment plant, vacuum filter and clarifier mechanisms etc.)	13,110	13,110
Pumping to and from treatment plant	———	17,500
Pumping to and from treatment plant and cooling	32,800	———
Sludge disposal	73,600	69,000
Operators	55,000	55,000
		$351,985
Oxygen plant 45 metric tons/day (50 tons/day)	66,000	
Power	66,000	
Supervision	2,500	
Maintenance	8,700	
Water	1,340	
	$362,400	
Credits		
Bleach plant 5 tons/day at $30	-52,500	
Excess to sell 5 tons/day at $19.30 80 percent of time	-27,020	
O_2 storage tank rent	-8,400	
	-87,920	
Total	$274,480	

Table C—40 presents a sample discounted cash flow computation for the pure oxygen system using capital and operating cost data from Tables C—36 and C—39, respectively.

TABLE C—40. SAMPLE COMPUTATION OF NET PRESENT VALUE AFTER TAX, CASH FLOW COST FOR PURE OXYGEN SYSTEM WITHOUT CLARIFIERS (DISCOUNT RATE OF EIGHT PERCENT)

Item	Amount
Capital Cost	
Portion eligible for 60 month amortization	$4,313,817
Portion eligible for 15 year DDB depreciation	75,000
Total capital cost	$4,388,817
Annual operating cost (excluding amortization and depreciation	275,000
Tax rate	50%
Investment tax credit, 7 percent on $75,000	$5,250
Net present value after tax cash flow cost	
Total capital cost	$4,388,817
Less: present value of annual incremental cash in flow	533,347
Cost	$3,855,470
Cash flow schedule	

Year	(1) Annual Operating Cost	(2) 60-Month Amortization	(3) 15-year DDB Depreciaton	(4) Incremental Cash Before Tax (1 + 2 + 3)
1				
2	$137,500	$431,382	$5,003	$573,885

Year	(5) Tax Savings 50% of (4)	(6) Incremental After Tax (4 — 5)	(7) Investment Credit	(8) Incremental Cash Flow-In (Out) (2 + 3 - 6 + 7)
1				
2	$286,943	$286,942	$5,250	$533,347
Total	$4262,500	$4,313,817	$75,000	$8,651,317
	$4,325,660	$4,325,657	$5,250	$68,410

Table C—41 shows the results of the same computation on the three systems considered, discounted at 8 percent and 10 percent.

Judging strictly from the net present values in Table C—41 at both discount rates the system choices would rank in the order (1) the oxygen system with rectangular secondary clarifiers, (2) aerated lagoons, and (3) oxygen with circular secondary clarifiers. However, important uncertainties led to further comparisons.

For example, Chesapeake knew of no rectangular clarifiers operating as satisfactorily as circular ones. Upon investigation this was confirmed but not before the rectangular clarifiers were retained for comparison as the oxygen system was broadened as described below. Furthermore, the second-choice aerated lagoons possessed the uncertain disadvantages already mentioned. Finally, the certainty of forthcoming standards for odor, or emission of odorous sulfur compounds, led Chesapeake to consider the relative costs of a black liquor oxidation system using pure oxygen or oxygen from air. Table C—42 shows the estimated capital costs for the systems considered previously and BLO with air versus BLO with pure oxygen.

TABLE C—41. COMPARISON OF NET PRESENT VALUE AFTER TAX, CASH FLOW COST (WASTE TREATMENT)

		Oxygen with—	
	Aerated lagoons	Circular	Rectangular
Capital costs	$3,571,583	$4,388,817	$3,993,447
Operating costs	352,000	275,000	275,000
Discounted at 8 percent	3,674,727	3,855,470	3,612,114
Discounted at 10 percent	3,576,184	3,815,251	3,562,945

TABLE C—42. CAPITAL COSTS OF ALTERNATE WASTE TREATMENT AND BLACK LIQUOR OXIDATION SYSTEMS

Item	Capital cost
Aerated lagoons	$3,571,583
Air BLO	450,000
Contingency at 25 percent	112,000
	$4,133,583
Oxygen with circular clarifiers	$4,388,817
Additional O2 plant	244,000
	$4,632,817
Oxygen with rectangular clarifiers	$3,993,477
Additional O2 plant	244,000
	$4,237,477

Table C—43 shows estimated annual operating costs for the systems listed in Table C—42 divided into oxygen applications and air applications.

TABLE C—43. OPERATING COSTS OF ALTERNATE WASTE TREATMENT AND BLO SYSTEMS

Item	Cost	
	O_2 applications	Air applications
Maintenance	$13,900	$14,300
Chemicals:		
Nutrients	45,500	59,500
Sludge conditioning	——	16,600
Chlorination	24,600	24,600
Power:		
Aerators	25,350	82,375
Pumping (within treatment plant, vacuum filter and clarifier mechanisms)	13,110	13,110
Pumping (to and from treatment plant)	——	17,500
Pumping (to and from treatment plant and cooling)	32,800	——
Sludge disposal	73,600	60,000
Operators	55,000	55,000
Oxygen plant (85 tons/day)		
Power	212,000	
Water	2,180	
Maintenance (1.5%)	12,000	
Supervision	2,500	
Air BLO		
Power		52,000
Maintenance (1.0%)		4,500
Supervision		5,000
Credits:		
Bleach, 5 tons/day at $30 ton	-52,500	
Liquid sell-out	-37,800	
Liquid sell-out, 1,960 tons/year at $19.30 ton	-37,800	
O_2 storage tank rent	-8,400	
	$313,840	$413,485

Notes:

(1) Neutralization costs not included.

(2) Cost of antifoam not included.

(3) Includes $FeCl_3$ addition at 5 percent of dry sludge weight.

Cost Basis:

(1) Maintenance: 1 percent of mechanical x 0.1 percent civil construction costs. For oxygen system $6,000 additional added for maintenance of UNOX aeration and instrumentation package.

(2) Chemical Costs

Ammonia	0.035/lb
H_3PO_4	$9.30/100 lb
Chlorine	0.85/lb
$FeCl_3$	4.00/100 lb

(3) Power Costs: 7 mills/kWhr or $43.70/hp-yr.

(4) Sludge disposal — landfill at $2.70/wet ton.

Using the capital costs data from Table C—42, the operating cost data from Table C—43 and the method displayed in Table C—40, the computed net present values of Table C—44 were obtained.

Again, as in Table C-41, the oxygen system with rectangular clarifiers appeared to be the best economic choice. However, in the meantime Chesapeake had not been able to establish operating reliability of the rectangular clarifier and rake mechanism. So, this system was abandoned.

Comparing the net present values of the oxygen system with circular secondary clarifiers and the aerated lagoons with air BLO, the former had a very slight economic edge at both discount rates. Ignoring this slight advantage, but giving considerable weight to (1) the disadvantages of the aerated lagoons, (2) the substantial space requirements of air BLO, and (3) the anticipated requirement to scrub the off-gases from air BLO, Chesapeake chose waste treatment and BLO with pure oxygen.

TABLE C—44. COMPARISON OF NET PRESENT VALUE AFTER TAX, CASH FLOW COSTS (WASTE TREATMENT AND BLACK LIQUOR OXIDATION)

Item	Aerated lagoons and air BLO	Oxygen with Oxygen BLO and Circular	Rectangular
Capital costs	$4,133,582	$4,632,817	$4,237,477
Operating costs	414,000	314,000	314,000
Discounted at 8 percent	4,280,688	4,168,098	3,925,759
Discounted at 10 percent	4,163,270	4,114,409	3,862,122

Approval of System by State Water Control Board—

After selecting the wastewater treatment system as described in the previous sections, Chesapeake applied (84) to the State Water Control Board for approval of the system. Information gathered on in-plant changes to reduce flow and pollutant loads and on the parametric limits on system operation was organized in accord with the special SWCB form prescribed.

Initial approval was quickly received (85) from the SWCB but certain conditions and terms, unacceptable to Chesapeake, were made a part of the approval. Chesapeake requested(86) revision or deletion of said conditions and terms as being unrealistic for the type of manufacturing operation and the type of waste treatment system planned. Agreement was reached with minimum delay and final approval of the system was received(87) from the SWCB in November 1972.

Process Design, Engineering and Installation

Conversion of the Hydroscience process design recommendations into a full-scale, engineered, and installed system was undertaken on a turn-key, competitive bid basis with three consulting and construction firms. Two of the firms submitted proposals, the third declined to bid. The choice between the two bidders was difficult since both featured approaches with appealing strong points. Both were based on identical information from Chesapeake and vendors who proposed to supply different parts of the WWTP. The documents supplied provided information(82) on the basic process design; on-site location and arrangement of equipment at Chesapeake's West Point, Virginia mill(88); on the bioreactor design and the oxygeneration system with the controls(89) which Chesapeake purchased directly from the vendor; on the detailed flowsheets and equipment specifications(90) for a 90.7 metric ton/day (100 ton/day) cryogenic oxygen plant, most of which Chesapeake purchased indirectly from the vendor through an oxygen supplier; on a liquid oxygen storage system(91) purchased directly by Chesapeake; and on site sub-soil characteristics(92) of the proposed site.

Chesapeake examined both bid proposals carefully, met several times with both parties and accepted Rust Engineering Proposal No. P272 originally dated July 7, 1972(93) but executed on November 15, 1972 by Rust and Chesapeake officers as Proposal No. 272-B, Revision 2 dated November 2, 1972. Correspondence covering negotiations and changes is located in certain Chesapeake engineering files.(94)

Chesapeake issued Purchase Order No. A8990 to Rust Engineering on February 2, 1973 to cover "lump sum engineering and construction of effluent treatment facilities". Construction started March 17, 1973, two weeks behind schedule because of excessive rainfall. Other factors such as an extended carpenters' strike and unavailability of carpenters after the strike delayed construction seriously. These delays were detailed(95) by Rust to Chesapeake from a daily construction diary. The original projection for startup was October 31, 1973. This date was delayed to December 31, 1973 then to March 1974. Reasons for the changes in projected startup were reported to the SWCB.(96)

Figure A—8 shows a schematic flowsheet of the WWTP. Figure A—9 depicts an engineering perspective drawing of equipment location and layout finally selected. Appendix A includes a detailed description of the major equipment installed in the WWTP and oxygen plant.

The primary part of the treatment system, including sludge dewatering began operation on April 23, 1974. Continuous operation of the oxygen plant began May 5, 1973, followed shortly by operation of the secondary treatment system on May 16, 1973. Problems encountered are described in a subsequent section.

In-Plant Effluent Collection System (IECS)

This system is reported in a separate section since the design, engineering, and installation were handled under different contracts with Rust Engineering.

Chesapeake issued Purchase Order No. A6285 dated June 27, 1972 to Rust to provide engineering and design service necessary for the installation of an In-Plant Effluent Collection System (IECS). The IECS was to be designed in accord with Rust Proposal No. P197, Revision No. 1 under Rust Contract No. 21—1203C, an extension of an existing Contract No. 21—1203 for engineering services to develop and implement an emissions control program for Chesapeake. The relation of this collection system to the overall waste treatment system is shown in Figure A—8. Figure A—7 shows the in-plant sumps of the collection system, including a fifth sump not covered by the contractual documents mentioned above. Design and engineering arrangements for the latter will be referenced below.

Chesapeake issued Purchase Order No. A11274 dated May 5, 1973 to cover installation of the IECS designed under Contract No. 21—1203C and developed further in Rust Proposal No. P377, Revision No. 2, dated March 28, 1973. Major design features of the IECS are included in Appendix A. Table C—45 details the effluent streams flowing to the initial four lift stations as well as those flowing to the more recent fifth lift station (for the kiln area).

TABLE C—45. FLOWS TO IN-PLANT EFFLUENT COLLECTION SYSTEM

Lift station (sump)	Effluent flows (from)
Paper mill	1. Unbleached linerboard and paper machines 2. Roll grinder room 3. Hydropulper in secondary fiber system
Pulp mill	1. Pulp washing 2. Screen room 3. Batch washer filtrate tanks 4. Black water tank overflows 5. Turpentine recovery system 6. Tall oil plant sump
Caustic	1. Nos. 3 and 4 recovery boilers 2. Batch blow tank 3. Causticizing area 4. Evaporator area
Bleach plant	1. Bleach plant—all stages 2. Chlorine dioxide generator system
Kiln area	Normal wastewater, overflows, and spills from: 1. Boiler feedwater plant 2. Power plant general 3. Maintenance and stores 4. Lime mud storage 5. Barn bin 6. Wet end of lime kiln 7. White liquor tank 8. New strong waste tank

The kiln area lift station (sump), shown as the last one in Table C—45, was installed to collect accidental spills and overflows to minimize discharge of chemicals to the Pamunkey River. Chesapeake issued Purchase Order No. C—0621, dated September 13, 1976, to Rust Engineering to cover "detailed engineering services for an additional effluent discharge control program to provide additional effluent collection capabilities to the existing system discharging to the Wastewater Treatment Plant." Rust Proposal No. 21—1773A dated August 24, 1976 contained details of the system which were expanded further in a Rust Final Scope of Work document(97). This document followed reports submitted earlier to Chesapeake in March and April 1976.

Chesapeake issued Purchase Order No. C—0965, dated January 5, 1977 to Tidewater Construction Corporation to "Provide Construction Services for Installation for an Effluent Discharge Control System." Construction started in May 1977 and the system was completed in January 1978. Figure A—7 shows how this lift station fits into the overall effluent collection system. Figure A—11 shows the shape and dimensions of the lift station. Appendix A, describes the major equipment associated with the lift station.

Wastewater Treatment System Evaluation

This section of the report presents Chesapeake's experience in starting up the WWTP, correcting mechanical and process problems (which were few) and WWTP performance data taken during and after the grant period. Also covered are steps taken after the grant period to control filamentous organism growth in the system.

Mechanical and Process Problems—

Perhaps the most troublesome problem of all was plugging of sump pumps at the lift stations; this story is detailed in the following paragraphs. Success and problems with each of the other subsystems are covered subsequently.

Lift station pump problems—Two vertical turbine pumps (which would pass 25.4 mm [1 in.] solids) were initially installed in the paper mill, pulp mill, and caustic area lift stations or sumps. These are described in Appendix A. Bar screens (also described in Appendix A) were installed in the paper mill and pulp mill sumps but not in the caustic area sump. Chesapeake assumed that the bar screens would remove solid material too large to pass through the pumps and that there would be no large solids coming to the caustic area sump. Both assumptions were wrong. Both pumps at all three lift stations plugged frequently and required excessive maintenance.

One of Chesapeake's construction progress reports(98) discussed the operation of other sump pumps serving paper mill, pulp mill, and caustic areas and recommended improvement to minimize plugging of pump suction by March 1, 1975.

This problem was a number one priority and was being investigated to the fullest extent possible. Corrective measures under investigation included pump back wash connections, inplant modifications to minimize plugging material, new pump impellers, and new pumps.

It became apparent that at least the caustic and paper mill lift pumps had to be replaced with pumps capable of passing solids. Quotations were obtained from pump manufacturers for replacement pumps capable of passing these solids. Every effort was made to minimize the plugging problem so that the lift stations could be operated until replacement pumps were installed.

A rigorous preventive maintenance program was established and level alarms were installed in each lift station to alert operating departments to the decreasing volume of wastewater being pumped from the station.

Turbine pumps for the bleach plant were ordered in 317 stainless steel at the same time (January 11, 1973) as the pumps for the lift stations mentioned above. However, they were long delivery items. The problem was explained in a letter to the SWCB(100).

The bleach plant effluent had not yet been treated on a continuous basis because of problems with the bleach plant sump. The lift pumps were ordered, of all 317 stainless steel, January 11, 1973. Delivery was promised on July 9, 1973. The pumps were delivered and events thereafter occurred as follows:

(1) Two discharge heads were received on January 30, 1974.

(2) Two pump bowl assemblies were received on September 13, 1974.

(3) Two oilers were received November 11, 1974.

(4) Two connector bearing and suction case plugs received were carbon not stainless steel.

(5) The discharge heads had to be reworked by Chesapeake (October 25, 1974) to meet the certified prints.

(6) Chesapeake made bearing sleeves for the pumps on October 22, 1974.

(7) Chesapeake substituted a quart size oiler because the right size oiler had not been received.

(8) One pump was started October 28, 1974, ran for three days and lost a bearing.

(9) A factory representative inspected the pumps on November 18, 1974.

Interspersed with the above happenings were efforts to pump the bleach plant effluent to the waste treatment plant. Each trial ended with the suction side of the vertical lift pump plugged.

In this same letter(100) Chesapeake requested an Extended Implementation Schedule under NPDES Permit No. VA-0003115 from December 31, 1974 to August 1, 1975. This was approved by letter(101) from the SWCB with the condition that periodic construction progress reports be submitted. Item 3 from the first such report(98) was quoted above on three sump pumps. Item 1 from the same report described the situation with the bleach plant sump.

Chesapeake met the requirement to have bleach plant sump pump operational by January 15, 1975. The bleach plant lift station was operational, however, there still were problems with plugging and bearing failures. The bearing failures were investigated by our consultants and the pump manufacturer. The principal cause of these failures has been determined as abrasion resulting from contact between the pumped fluid and the bearing. A mechanical seal arrangement with fresh water seal water was devised and installed as soon as the mechanical seal was received. A special gland and packing arrangement with fresh water seal water was installed as a temporary solution until the mechanical seal was received.

The bleach plant pump plugging problem is similar to the paper mill, caustic, and pulp mill lift pump plugging problems.

Seeking some solution to the sump pump plugging problem Chesapeake sought information on available pumps and selected one for each sump that would pass solids up to 7 mm (2¾ in.) in diameter. Four pumps were quoted on May 19, 1975 with estimated delivery of 52 weeks. The pumps were ordered on Purchase Order no. A-27052, dated June 6, 1975 and were delivered July 1976.

Thereupon, Chesapeake fulfilled its commitment to the SWCB by installing the pumps with minimum, practical sump overflow. The bleach plant was down when the pump was installed about September 20, 1976 in the bleach plant lift station. The other three lift stations were out of operation for 78 hours (paper mill), 80 hours (pulp mill), and 81 hours (caustic) at different times October 5 to 10, 12 to 15, and 19 to 22, respectively. The outages were reported to the SWCB.

After installation of the new pumps, lift station performance improved greatly. One of the initial pumps was left intact in each lift station as a spare. The lift stations are presently (1978) operating well without unusual maintenance as before. In other words, the results expected were achieved.

Primary clarifier system—In general, the primary clarifier system has worked well. The notable exceptions are (a) plugging of the sludge piping, (b) skimming and handling the floatable materials, and (c) very occasional rake drive malfunctioning.

Sludge line plugging was stopped by replacing butterfly control and block valves with V-ball and gate valves, respectively. Also, all check valves were removed from the sludge lines.

The scum skimming and disposal problems was worsened by an increase in secondary fiber repulped, cleaned, and used in Chesapeake's products. Plastic film, styrofoam, and other floating materials could be skimmed from the primary clarifier surface but scum lines and pumps plugged frequently. The problem was solved by providing a perforated dumpster box to receive the skimmed material. Water drained freely to the original scum tank and could be pumped back to the primary clarifier. The plastics, etc., were hauled to landfill.

The torque sensor on the rake mechanism drive was set to alarm at 48 percent of design. At 88 percent of design torque, the drive was designed to shut down automatically. Ocassionally, the torque climbed toward the first alarm set point. In such cases, the wastage rate of primary underflow was increased or an inquiry launched to find the mill source of heavier fiber or other settleable solids. One or both steps usually corrected the problem.

The most serious problem with the primary clarifier system occurred almost two years after the grant period but it is worthy of mention here. The problem is best described in the report(102) of such events to the Regional Office of the SWCB.

The report stated that we had to shutdown our primary clarifier due to two problems. The following description of the clarifier drive assembly is necessary for an understanding of the clarifier problems.

The drive assembly consists of a 1.5-m (60 in.) diameter heat treated cast steel main spur gear mounted on precision bearings, two secondary gear reducers driving the main spur gear 180° apart, two primary worm gear reducers that drive the secondary gears, and worm shaft mounted cone reducers that drive the primary reducer. The two motor balanced type drive head was designed for 280,000 ft-lb of torque. A mechanical torque overload switch that is dependent on movement of the

No. 1 drive primary reducer worm shaft prevents overloading the rake arms and drive assembly. The rake arms are supported by the main spur gear housing.

On September 18, 1977 at approximately 7:00 a.m., the primary clarifier shutdown due to high torque. An investigation indicated there was not sufficient fiber in the primary clarifier for a torque overload. Further investigation revealed the No. 2 drive cone reducer had failed. A new cone reducer was installed and the primary clarifier started up by 6:00 p.m. on September 18, 1977. During the investigation and reducer replacement, the primary clarifier was by-passed. All wastewater except the paper mill effluent was treated in the secondary biological treatment system. The paper mill lift station was shutdown (our Cooling tower and secondary clarifiers cannot handle the paper mill effluent fiber) and overflowed to the river through discharge 010.

At 10:00 p.m. on September 18, 1977, the primary clarifier was again shutdown. This shutdown was caused by the complete failure of the main spur gear housing (the housing sheared at the bearing raceway, dropping all 119 of the ball bearings into the clarifier). Again we by-passed the primary clarifier and shutdown the paper mill lift station.

On September 19, 1977, the clarifier manufacturer's service representative inspected the clarifier damage. The service representative indicated they had never had a similar failure of the main spur gear housing.

A new main spur gear housing with spur gear and ball bearings was shipped on the afternoon of September 20, 1977 and received on the morning of September 23, 1977 (note: a direct through truck with two drivers was used to transport the housing from Salt Lake City, Utah to West Point, Virginia).

Our investigation of the cause of the main spur gear housing failure indicates the omission of two limit switches, one on each of the two primary reducer worm gears, was the cause of the failure. These limit switches shut down the clarifier if there is a mechanical failure in either of the cone drives, primary reducers, or secondary reducers. Without these limit switches, a mechanical failure in one of the two balance head drive system results in the full torque being transmitted to the main spur gear through the active secondary gear. An unbalanced force results. We speculate that the failure of the No. 2 cone drive created this unbalanced force with a resulting main gear housing failure.

To prevent a recurrence we have installed the above two limit switches. In addition, during this brief clarifier outage we overhauled the two drive assembly systems on the clarifier.

Contrary to some expectations, there have been no consistent or persistent foam problems in the primary clarifier system. When a black liquor spill was not or could not be diverted to the strong waste tank, a temporary foam problem developed. The foam dissipated as the black liquor passed on through the WWTP.

Nutrient addition system—The initial and replacement nutrient addition systems are described in Appendix A.

The initial vaporizer and flow control system installed on the liquid anhydrous ammonia storage tank did not work well consistently. The principal problem was control of the water flow and pressure to the eductor which drew the gaseous ammonia from, and, therefore, the liquid ammonia into the steam vaporizer.

The replacement system with a rotameter, needle control valve, and long-pipe ambient vaporizer has generally operated well. The first rotameter has been replaced with a smaller one which, at the desired ammonia addition rate, places the float in a position on the scale much easier to read.

The ammonia addition rate has consistently been about 675 kg/day (1500 lb/day). The WWTP operators check this addition rate periodically by the change in level of the horizontal cylindrical storage tank. A special calibration chart showing volume versus vertical depth was prepared for the WWTP operators.

Another check on adequacy of ammonia addition rate was measurement of residual ammonia in the treated effluent. There were and are still uncertainties in the analytical method used. Chesapeake has sought counsel from the National Council of the Paper Industry for Air and Stream Improvement (NCASI) and recently have undertaken laboratory studies in-house. Results are inconclusive thus far.

There have been no serious probems with the phosphoric acid storage and addition system. The positive displacement metering pumps described in Appendix A have worked well. The addition rate, as 35 percent phosphoric acid, has been about 1350 kg/day (3000 lb/day). Nutrient additions expressed as BOD removed: nitrogen: phosphorus have been about 100:4:1.

Several sources of phosphorus, as phosphoric acid, were tried in an effort to reduce nutrient costs. Some that were tried (less expensive than the one selected and now being used) did not seem to work well in the WWTP system. The reasons were not sought since other sources of phosphoric acid were available. Initially, Chesapeake used an expensive 75 percent phosphoric acid but has been using a less expensive 35 percent phosphoric acid for almost two years.

Defoamer storage and addition system—The only changes made in the original system design for defoamer storage and addition were (a) adding a second storage tank to permit use of two different types of defoamers, (b) adding agitators, and (c) providing either steam or electric heaters to keep the storage tank contents liquid.

The defoamers were added, one ahead of the cooling tower and the second after the bioreactor. Additive rates varied depending on need for foam control.

A number of defoamers were tried until two that appeared most effective were found. There is little point in identifying these particular defoamers, since they probably would not perform well in a different application. The general properties permitted use of one at higher pH levels, the other at lower pH levels.

Cooling tower—Operation of the cooling tower has been very satisfactory. The metering orifices/diffusers in the distribution box at the top of the tower plugged occasionally but could be manually cleaned easily. Some slime occurred in the tower fill but this was minimal and caused no serious problem. Foaming and plume drift were expected in the cooling tower but neither have been a problem even though the tower is only 91 m (300 ft) from the nearest residence.

Until the scum skimming system on the primary clarifier was improved, as mentioned earlier, floatable plastic pieces, styrofoam, etc., carried over the primary launder and eventually went to the cooling tower cold well. Manual removal was a toilsome task. This is still a problem to some degree.

Bioreactor—The dimensions and general arrangement of the bioreactor system are given in Appendix A. This system is the heart of the WWTP, where a high percentage of the total BOD removed from the waste stream occurs.

During startup of this system, which is covered to maintain a gas phase at low positive pressure (100 to 120 mm water maximum) and rich in oxygen, it was difficult to locate and eliminate all gas leaks. It was imperative to do so and was finally achieved after several weeks of searching. The only other disadvantage of the covered system was being unable to view any problems that might occur inside the reactor. Fortunately, there have been few, in fact, almost none. During a black liquor spill soon after startup, the aerator motors overloaded for reasons unknown. Heavy foam in the gas space was suspected. Further study showed the foam increased the level in the reactors by partially blocking flow between the reactor to the secondary clarifiers. The addition of defoamer at the splitter box ahead of the bioreactor cured the problem. Since then it has been standard practice to feed defoamer at this point, when a black liquor spill reaches the WWTP, in addition to the defoamer fed ahead of the cooling tower. The problem has not recurred.

It is important to maintain an adequate supply of high-purity oxygen to assure high oxygen partial pressure in the gas space. There have been no significant problems with the system controlling oxygen feed to each of the two first stages. Static pressure sensors and control valves admit oxygen if the slight positive gas pressure drops below the set point.

The safety system consists of a hydrocarbon analyzer and air purge blower; it has operated well. At startup, and for sometime thereafter, the set point of the hydrocarbon analyzer was mistakenly adjusted to 12.5 percent (one half of the design set point) of the lower explosive limit (LEL) for oxygen flow cut off and air purge blower startup. This caused inordinate problems at the WWTP as explained in a letter (99) to the SWCB.

The letter reported that one of the major unexpected problems was due to turpentine spills.

One consequence of the spills affected the high MLVSS concentration required to attain the design F/M ratio. With frequent turpentine spills that reduce the MLVSS concentration (this occurs because turpentine trips the hydrocarbon analyser set-point, resulting in the automatic shutoff of feed oxygen to the reactor and the automatic purging of the reactor with air) and with poor sludge thickening in the secondary clarifier resulting from filamentous organisms, we were not able to achieve our design F/M.

To correct the turpentine problem we installed a conductivity probe with automatic control valves to route turpentine and black liquor carryovers to the evaporator boilout tank, where the spills are returned to the recovery process. We also increased the setpoint on the UNOX hydrocarbon analyser to the level recommended by Union Carbide Corporation. At the new setpoint (which is double the percentage of the lower explosive limit of our initial setpoint. we are able to operate during all except a major spill. In fact, we have seen only one turpentine spill thus far that would have exceeded the new setpoint (25 percent of the LEL).

When any hydrocarbon concentration in the gas phase reaches 50 percent of the LEL the aerators automatically stop. Since startup there have been no shutdowns of aerators for this reason. The hydrocarbon analyzer is located in the control room and receives a gas sample from the bioreactor every ten min.

Other safety features of the bioreactor include (a) no installation of combustible materials in the gas space, (b) aerator drive motors located completely on top of the reactor cover outside the gas space, (c) aerator drive shafts operating through water seals, and (d) pressure vacuum seals on all stages. The water seals have a 483 mm (19 in.) column. The pressure-vacuum seals are set to release pressure at 152 mm (6 in.).

To maintain adequate oxygen partial pressure in the gas space of all stages for oxygen transfer to the liquid phase, a small volume of gas is vented from the last stage at 35 to 50 percent oxygen content. The rest of the vent gas is essentially carbon dioxide. When the alkalinity of the phase is low, carbon dioxide generated by biodegradation of the organics is not held in solution to the extent it is when the alkalinity is normal or high. Thus, the carbon dioxide partial pressure in the gas increases substantially. To offset such an increase it is necessary to increase oxygen feed rate above the normal oxygen consumption rate until oxygen partial pressure in the vent gas is adequate to ensure required oxygen transfer. In this situation, which has occurred infrequently, there is some oxygen tonnage or blow through that is not transferred to the liquid phase. Such an event manifested itself as oxygen consumption higher than 1.2 to 1.3 kg/kg BOD (1.2 to 1.3 lb/lb BOD) removed or in an oxygen utilization efficiency of less than the normal 90 to 92 percent.

Chesapeake was concerned about mixing in the bioreactor or the accumulation of solids in dead spots in the large reactor chamber. The bioreactor was not shutdown expressly for an inspection but an unexpected series of events created an opportunity to open the reactor. The events were described by letter (103) to the SWCB, which covered two incidents that occurred that had a bearing on the compliance terms of NPDES Permit No. VA0003115 by August 1, 1975.

On the night of April 3, (1975) at approximately 8:00 p.m., heavy winds blew down the up-draft tube on the third stage of the bleach plant. This caused a complete shutdown of the bleach plant and, therefore, eliminated the bleach plant effluent from the waste stream going to the Wastewater Treatment Plant.

The second event took place on April 6 when, by an operator error, a strong waste stream with high turpentine content was pumped to the strong waste tank instead of being returned to process. It was pumped then from the strong waste tank to the treatment plant and immediately caused an automatic cutoff of the oxygen supply and purging of the bioreactor with air. This second event, following so closely the first described above, apparently dealt a lethal shock to bacterial population in the bioreactor. In spite of diligent attempts to batch feed the greatly reduced bacterial population to encourage their regrowth, the population continued to decrease to the point that the system was almost totally ineffective.

After such efforts to restore the system to the high levels of efficiency we were observing prior to these two events, our best judgment was to shut the treatment plant down, drain it and clean it completely, and reseed the bioreactor for a fresh start.

These happenings were particularly frustrating to us since we had achieved superior operation of the treatment plant the week before the events occurred.

Before cleaning the bioreactor as stated above, inspection showed little if any solids accumulation; this indicated good mixing and circulation in each stage. A small solids build-up around the interstage gas ports was easy to remove.

As already mentioned, sufficient oxygen must be fed to the bioreactor to maintain a partial pressure that assures transfer to the liquid phase. In so doing, the dissolved oxygen (DO) levels reached in each stage were (and are) higher than those observed in a conventional air activated sludge system. Typical DO levels by stage at design BOD_5 influent levels are 7 mg/L (1st stage) 9 mg/L (2nd stage), and 12 mg/L (3rd stage). Unusual DO levels in an air system are 0.5 to 1.5 mg/L.

Secondary clarifier system—The secondary clarifiers and sludge withdrawal mechanism are described in Appendix A. In addition to that description, this section describes the piping system between the bioreactor and the two clarifiers (see Figure C—26). Certain of the mechanical and operating problems which occurred were speculatively judged to be influenced by this piping system (such as the foaming problem described below—letter to the SWCB.(105)

A foam problem occurred in the secondary clarifiers. It varied in severity depending on the mixture of wastewaters going to the treatment plant. Since these were so variable to date, the reason for the foam problem was simply not apparent. It was not corrected with certainty until all waste flows were going to the treatment plant at a steady rate.

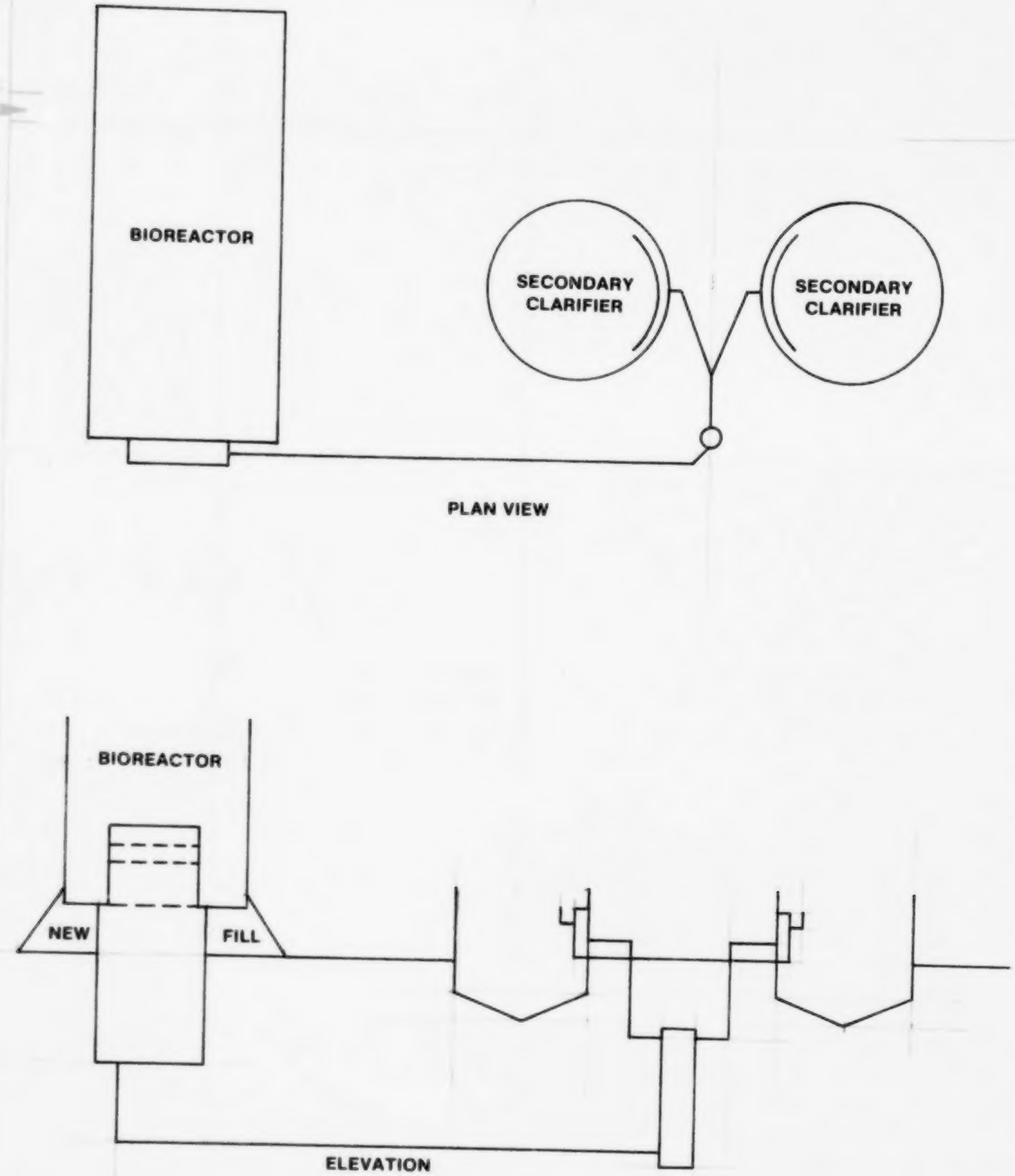

Figure C—26. Views of piping between bioreactor and secondary clarifiers.

Biodegraded effluent from the bioreactor flows over a weir, then by gravity to the secondary clarifiers. This is an excellent point for air entrainment. The elevation difference between the free surface after the bioreactor weir and the free surface of the peripheral feed raceway of the secondary clarifiers is about 0.49 m (1.6 ft). However, the bottom of the 0.92 m (36 in.) diameter pipe drops to an elevation underground of 6.48 m (21.25 ft) over a length of more than 106.8 m (350 ft). This means that the bioreactor effluent. including sludge recycle of 36 percent, flows turbulently under a pressure of almost 6.1 m (20 ft) of water or 59.7 kPa (8.7 psig) for about 1.6 minutes.

Recognizing that the following paragraphs represent oversimplification, it is possible, nevertheless, to establish an order-of-magnitude effect of carbon dioxide carried to, and released at, the secondary clarifiers.

For example, at a pressure of 60 kPa and temperature of 37°C (99°F), the equilibrium concentration of carbon dioxide in water is 0.167 g CO_2/100 g H_2O. At atmospheric pressure and the same temperature it is 0.104 g/100 gH_2O. Because minute gas bubbles are suspended, trapped, and compressed in the liquid phase flowing in the subterranean pipe, additional CO_2 could dissolve in the liquid phase to the extent of 0.063 g CO_2/100 g H_2O at equilibrium at 37° C (99° F). As the hydrostatic pressure goes down in the upflowing sections of the feed pipes to the secondary clarifiers, CO_2 (and O_2) will evolve and rise to the raceway free surface. Without sufficient free surface area and time for gas evolution and dispersion, the foam forming tendencies increase. Considering that it appears possible to release in excess of 8.5 m^3/min (300 cfm) of gas in each clarifier, it is surprising there are not constant foam problems.

Early design of secondary clarifiers being considered included a stilling box at the inlet point that would slow inlet velocity to the peripheral feed raceway and allow time for gas evolution and dispersion to occur. Because of very high additional projected costs, this design was adandoned and the inlet pipe entered the raceway directly at a 90° angle. This created a highly turbulent zone in the peripheral raceway and increased vulnerability to foam formation and overflow to the clarified effluent launder as well as over the outer walls of the clarifier.

It is a moot point whether or not a stilling feedwell would have reduced the severity of the foam problem. It is possible, however, to make a few comments on the foam problem itself. First, even after getting all flows to the WWTP, the occurrence of foam problems was not always understood or predictable. Second, as would be expected when a black liquor spill reached the WWTP, less than complete removal in the bioreactor caused a persistent and tenacious foam to form at the feed raceway of the secondary clarifiers; no amount of defoamer added after the bioreactor solved the problem. Third, water sprays which appeared effective when tried experimentally, did not solve the problem when installed around the entire periphery; eventually they were removed. Fourth, to contain the foam more effectively, runs of brick were installed, 280 mm (11 in.) and 330 mm (13 in.), respectively on the outer clarifier and clarified effluent launder walls. Thus far, the higher walls have been helpful but by no means have solved the foam problem. Chesapeake has no recommendations to offer as a certain solution to the problem except possibly a large, free-surface stilling well between the bioreactor and secondary clarifiers.

Three other secondary clarifier problems were: (a) short circulating of non-clarified effluent from the feed point to the clarified effluent launder, (b) starving of the peripheral feed raceway approximately one-third the circumference of the clarifier opposite the feed point, and (c) growth of filamentous organisms in the bacterial population. Fortunately, the first two problems could be solved by the same corrective step. This will be discussed first. The third problem was more complex but a solution was found as indicated below.

Regarding the first two problems, the bottom of the peripheral feed raceway was fitted with 76 mm (3 in.) diameter posts, or orifices, spaced 1.22 m (4 ft) apart. Likewise, the width of the raceway narrowed from 380 mm (15 in.) at the feed point to 230 mm (9 in.) diametrically opposite. The purpose was to achieve good distribution of feed around the entire periphery but this did not happen; the narrow part of the feed raceway periodically ran dry and accumulated solids which were difficult to remove. Also, there was visual evidence of short-circuiting near the feed point with excessive solids carryover into the effluent launder.

To correct both situations, seven orifices on each side of the feed point in both clarifiers were blocked with conical wood plugs. This stopped the short-circuiting and fed more bioreactor effluent to the narrow part of the raceway. Occasionally, at low flows, solids still accumulate at that point. Their removal manually with shovels is more difficult now since small pieces of plastic film, from the secondary fiber cleaning system, appear to bond the dried biomass tightly together.

During normal operation of the secondary clarifier system, a sludge blanket depth of about 1.53 m (5 ft) is maintained. This depth provides proper sludge aging at normal wasting and recycle rates and also enables good effluent clarification. Unfortunately, it has not been possible to maintain this blanket depth at all times. Also, the underflow solids concentration has not consistently reached the design level of 2.5 percent which was achieved and exceeded easily in the pilot plant work. This system being a highly complex biodynamic one, lack of conrol of blanket depth or underflow solids can be caused by numerous variants in the system. However, the presence of filamentous organisms in the bacterial population can strongly influence these two performance factors. Such organisms were observed under the microscope, the frequency depending on the nature of WWTP upsets due to spills of black liquor, turpentine or both; lack of nutrients; changes in F/M and changes in pH. Once the filamentous population reached a level considered intolerable because of excessive sludge blanket depth, low underflow solids, suspended solids carryover (TSS) and drop in MLSS, Chesapeake sought a method of control or selective elimination of such organisms. Gilbreath (105) summarized the influence of filamentous organisms on operating characteristics of the WWTP as shown in Table C—46.

A summary description of the first steps taken to control filamentous organisms was detailed in an internal memorandum (106) and was presented in a short paper (107) prepared for NCASI conference.

The filamentous organisms developed in our high-purity oxygen activated sludge wastewater treatment system through a combination of several of the classical causes of filamentous growth. This included insufficient nutrients, excessive organic load, and a low dissolved oxygen concentration. A literature review indicated that hydrogen peroxide had proven successful in the control of filamentous bulking without extensive damage to the desirable spherical bacteria.

A pilot plant study was conducted to evaluate hydrogen peroxide for control of filamentous organisms in a high-purity oxygen activated sludge system. Based on the results of the pilot-plant study, our full-scale UNOX waste treatment plant was treated with hydrogen peroxide.

Sphaerotilus, Escherichia coli, Geotrechum candidum, and Thiothrix are the organisms commonly identified as filamentous(108)(109)(110). These organisms are characterized by a high area-to-volume ratio; consequently, they have a metabolic advantage in an activated sludge reactor when compared to spherical aerobes(110). The filamentous organisms can therefore compete more readily for dissolved oxygen and nutrient when these concentrations are low.

Chudoba(108) has reported that complete mixed activated sludge basins with large dispersion numbers encourage filamentous growth. The filamentous growth can be controlled by aeration in multiple stages with a higher concentration gradient of substrate through the system(111).

Pasveer(109) reports that E. coli grows as a filamentous bacterium under acid conditions and is stimulated by the presence of glucose and CO_2. Nitrification gives the biological floc a low acidity and together with available nutrients produces an environment favorable to the growth of E. coli.

Other causes of filamentous growth are a high carbohydrate content in the feed, improper solids loading, or F/M, and shock organic loads.

Hydrogen peroxide has recently proven effective for the control of filamentous bulking. Keller(112) reports that dosages greater than 0.1 lb H_2O_2/lb MLSS under aeration reduces filamentous growth. Based on this work and other reports of hydrogen peroxide controlling filamentous growth(110), a pilot-plant study was conducted at Chesapeake Corporation.

A small pilot-plant model of our UNOX Wastewater Treatment System was operated on UNOX influent wastewater. A feed rate to the pilot-plant of 600 mL/min was selected to duplicate the

TABLE C—46. EFFECT OF FILAMENTOUS ORGANISMS ON THE WWTP

	Operating characteristics	
Parameter	**Without filamentous organisms present**	**With filamentous organisms present**
BOD_5	92 to 94 percent removal 30 mg/L avg effluent	90 to 94 percent removal 32 mg/L avg effluent
TSS	40 mg/L avg effluent	75 mg/L avg effluent
Color	15 to 20 percent removal	15 to 20 percent removal
Oxygen usage 90 percent oxygen utilization	0.6 kg O_2/kg BOD_5 (1.33 lb O_2 lb O_2/lb BOD_5)	0.6 kg O_2/kg BOD_5 (1.33 lb O_2/lb BOD_5)
MSVSS (95 percent volatile)	4,500 mg/L	4,000 mg/L
Food/micro-organism	$\frac{0.36\ \text{kg BOD}_5\text{/day}}{\text{kg MLVSS}}$ $\frac{(0.8\ \text{lb BOD}_5\text{/day})}{(\text{lb MLVSS})}$	$\frac{0.41\ \text{kg BOD}_5\text{/day}}{\text{kg MLVSS}}$ $\frac{(0.9\ \text{lb BOD}_5\text{/day})}{(\text{lb MLVSS})}$
Solids production	0.16 kg VSS/kg BOD_5 (0.35 lb VSS/lb BOD_5)	0.16 kg VSS/kg BOD_5 (0.35 lb VSS/lb BOD_5)
Sludge volume index	50	90
Initial settling velocity	1.68 m/hr (5.5 ft/hr)	1.07 m/hr (3.5 ft/hr)
Filter loading	34.2 kg/hr/m^2 (7.0 lb/hr/ft^2)	22 kg/hr/m^2 (4.5 lb/hr/ft^2)

hydraulic detention time of the full-scale system. Biological sludge from the full-scale unit was used to seed the pilot plant with filamentous organisms.

After a two-week pilot plant acclimation period, hydrogen peroxide was added to the return biological sludge at a rate of 40 mg/L based on the influent flow (0.13 kg of H_2O_2/day/kg MLSS under aeration). Figure C—27 shows the effect of the peroxide on the initial settling velocity (ISV), while Figure C—28 shows the effect on the sludge volume index (SVI). As shown, the one day dosage at 40 mg/L reduced the SVI from an average of 70 mL/g to an average of approximately 47 mL/g. However, there was very little effect on the ISV.

At a dosage of 80 mg/L (based on influent flow) as shown in Figures C—29 and C—30, the ISV increased from 7. to 16 ft/hr while the SVI decreased from 63 to 45 mL/g. Very little reversion of these settling characteristics occurred at the 80 mg/L H_2O_2 dosage.

Based on the encouraging results of the pilot plant study, a full scale treatment with hydrogen peroxide was planned.

In May 1976, operational problems with our wastewater treatment system dictated that hydrogen peroxide treatment be immediately initiated. The sludge blanket depth in our secondary clarifiers had risen to 12 feet (14 ft SWD clarifiers) with a SVI fluctuating between 70 and 100 mL/g and effluent TSS concentration of up to 200 mg/L.

Hydrogen peroxide (50 percent) was fed to the return biological sludge at 40 mg/L (based on the influent flow) the first day and at 80 mg/L for two additional days. Figures C—31 and C—32 show the effects on SVI and ISV.

As shown in Figure C—32, the SVI continued to decrease even after the termination of peroxide addition. The SVI leveled off at approximately 55 mL/g. The ISV did not change appreciably due to an increase in the MLSS concentration. As the SVI decreased, the secondary clarifier underflow concentration increased from 11,600 to 17,600 mg/L. As a result the MLSS increased from 3,800 to 4,750 mg/L.

The sludge blanket depth dropped from 3.66 m (12 ft) to 2.44 m (8 ft) during the peroxide addition. Effluent suspended solids averaged 120 mg/L during the treatment period.

At the conclusion of the peroxide addition, the filamentous population was sufficiently reduced for effective biological treatment. Even though the filamentous organisms were not completely eliminated, the system has not reverted to the pretreatment filamentous population levels.

It should be pointed out here that a dosage of 100 mg/L of H_2O_2 was also tried in the pilot-plant study reported by Gilbreath. Chesapeake wished to confirm that an accidental dosage at the 100 mg/L level would not upset the biological system. Actually, the changes in SVI and ISV were similar to those observed in the pilot-plant at a dosage of 80 mg/L.

Another approach (113) was used in October and November 1977 after some growth in filamentous population was observed microscopically. Although the operational problems ascribed earlier to filamentous growth had not yet developed to the intolerable level of April 1976, a different concept of control appeared trialworthy.

For example, bacteriological analyses showed the principal filamentous organisms to be either Thiothrix or Beggiatoa. Both use sulfides as a food source. Sulfides occur frequently in the WWTP influent and would be particularly high during green or black liquor spills. Oxidizing the sulfide content to zero, or some low level should, theoretically, starve these two organisms and others using

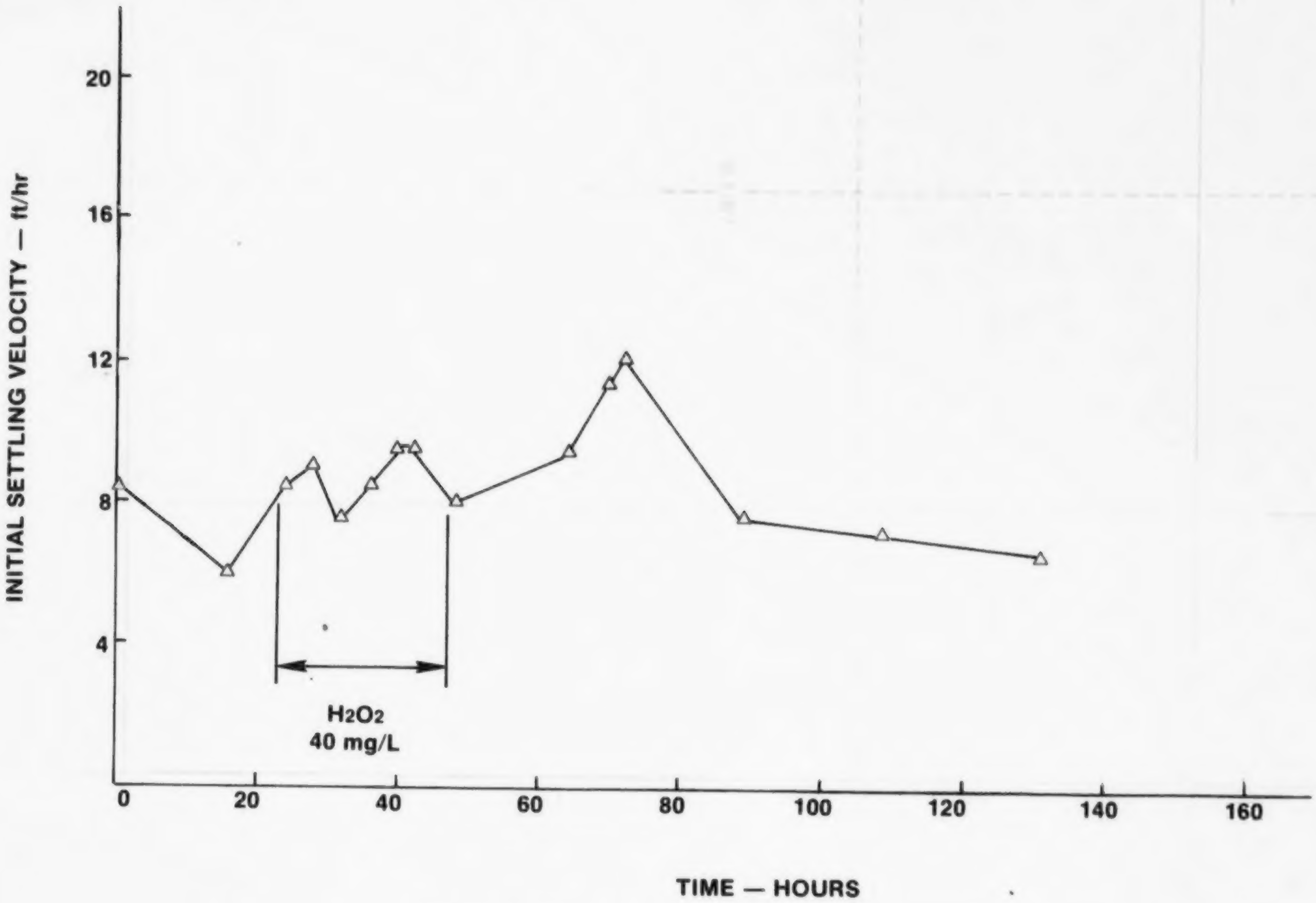

Figure C—27. Initial settling velocity observed in pilot-plant H_2O_2 treatment (40 mg/L).

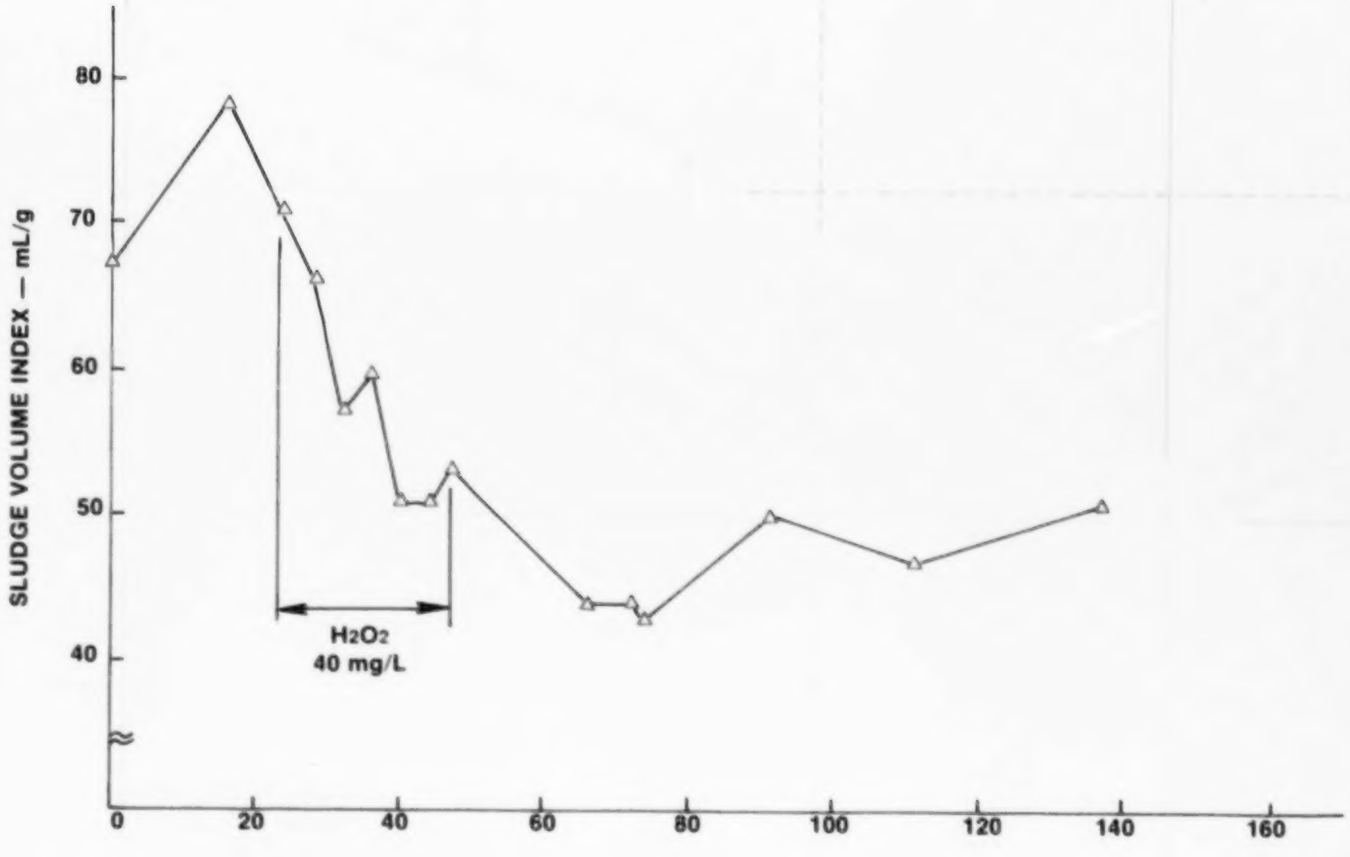

Figure C—28. Sludge volume index observed in pilot-plant H_2O_2 treatment (40 mg/L).

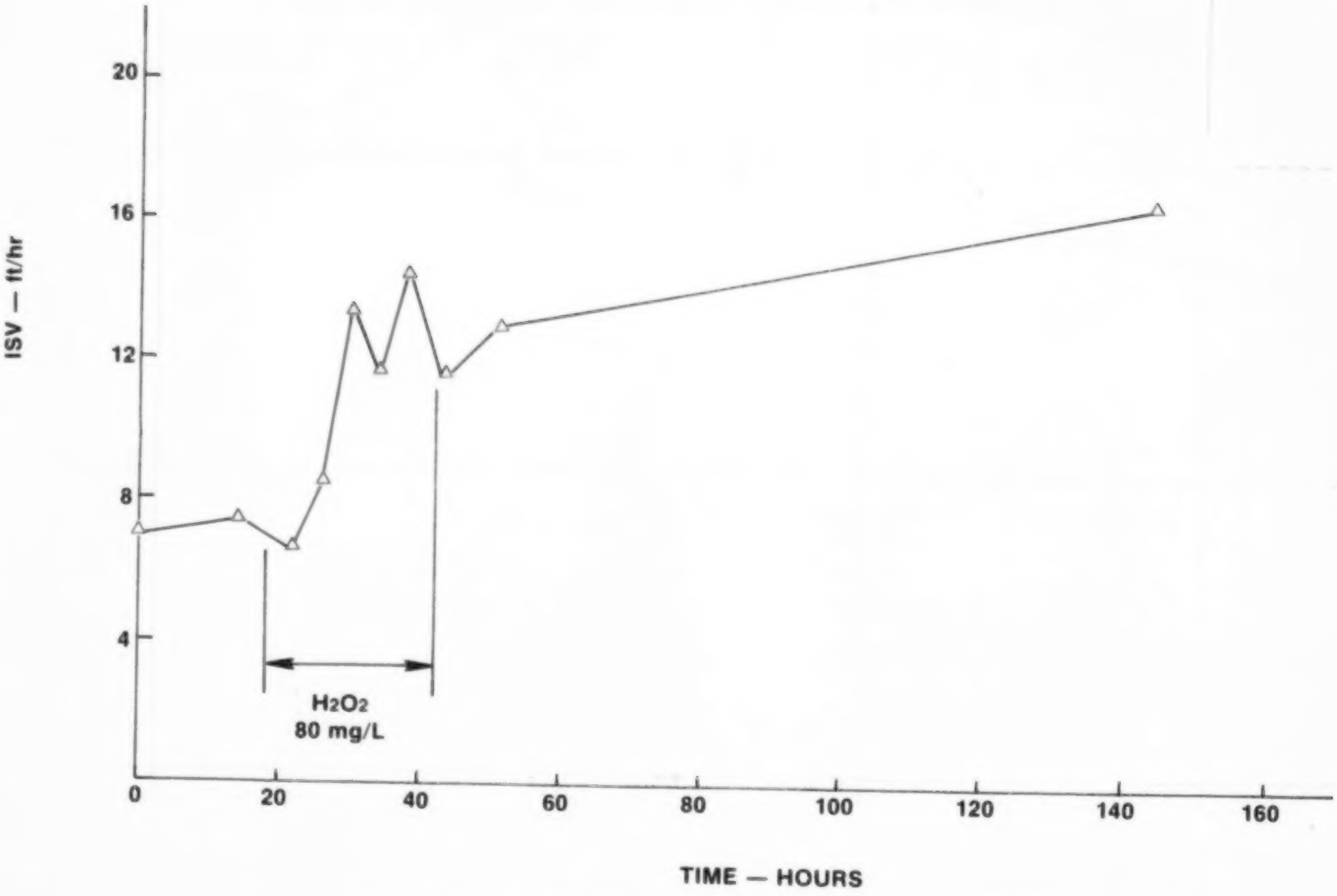

Figure C—29. Initial settling velocity observed in pilot-plant H_2O_2 treatment (80 mg/L).

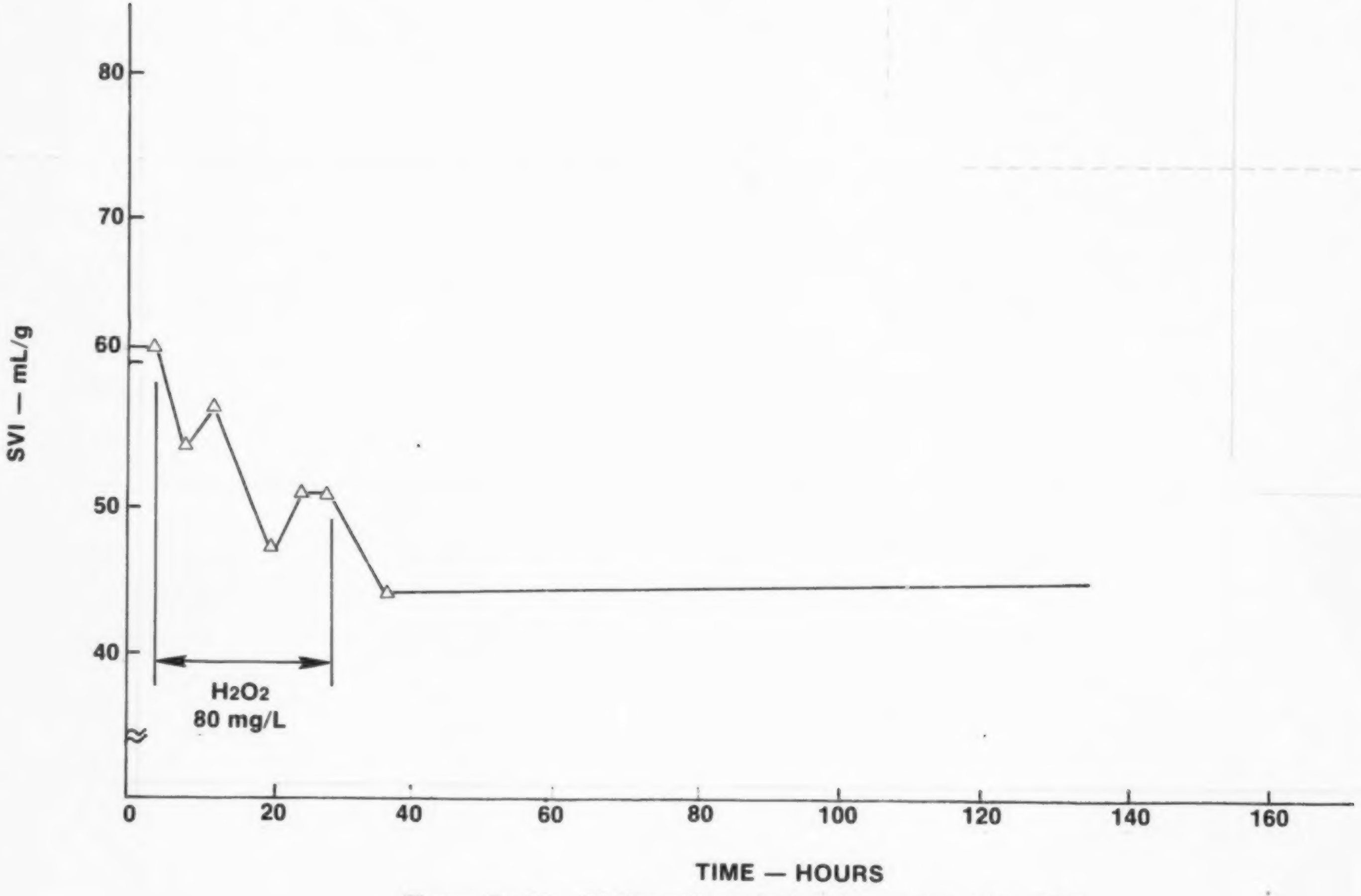

Figure C—30. Sludge volume index observed in pilot-plant H_2O_2 treatment (80 mg/L).

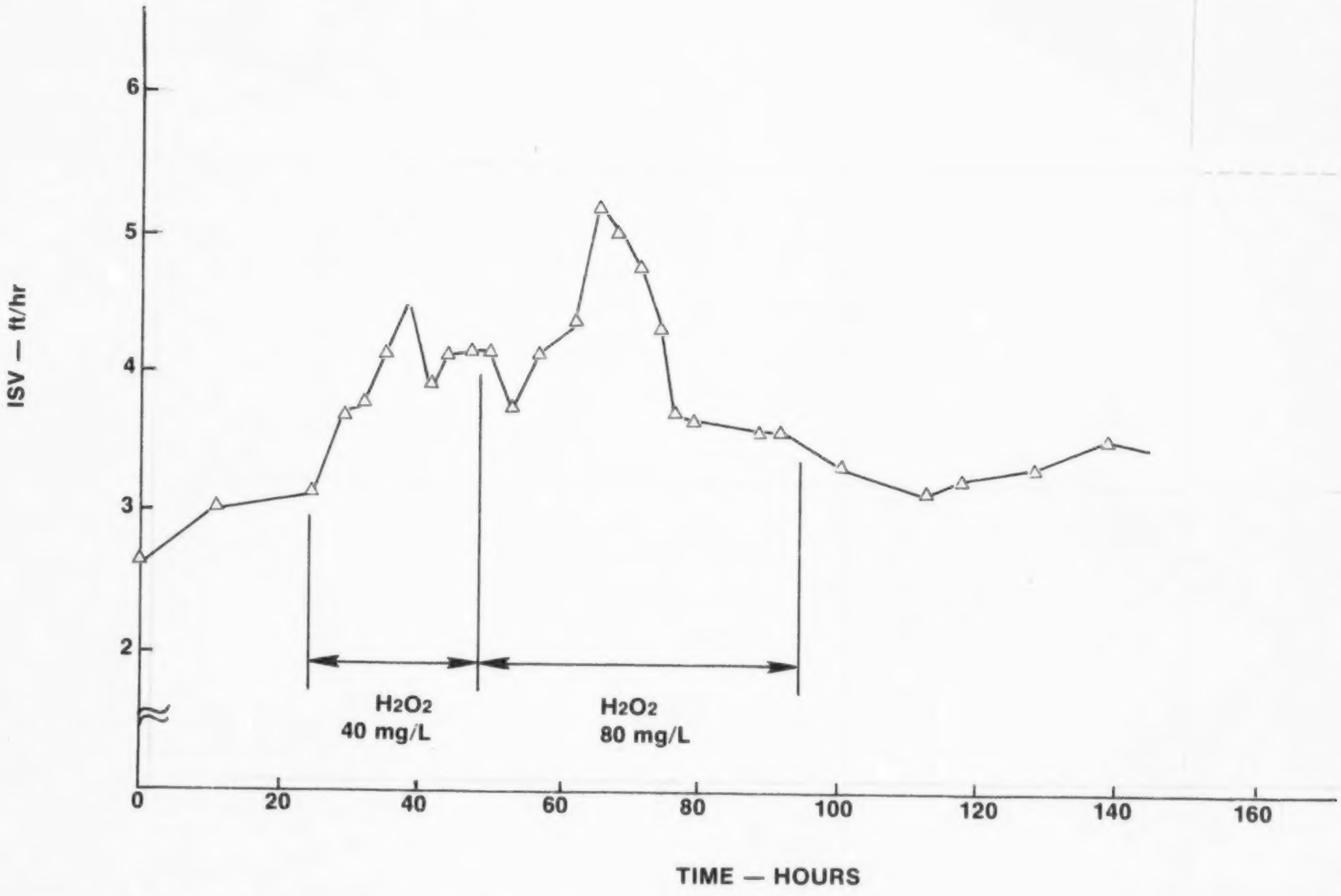

Figure C—31. Initial settling velocity observed in full-scale H_2O_2 treatment.

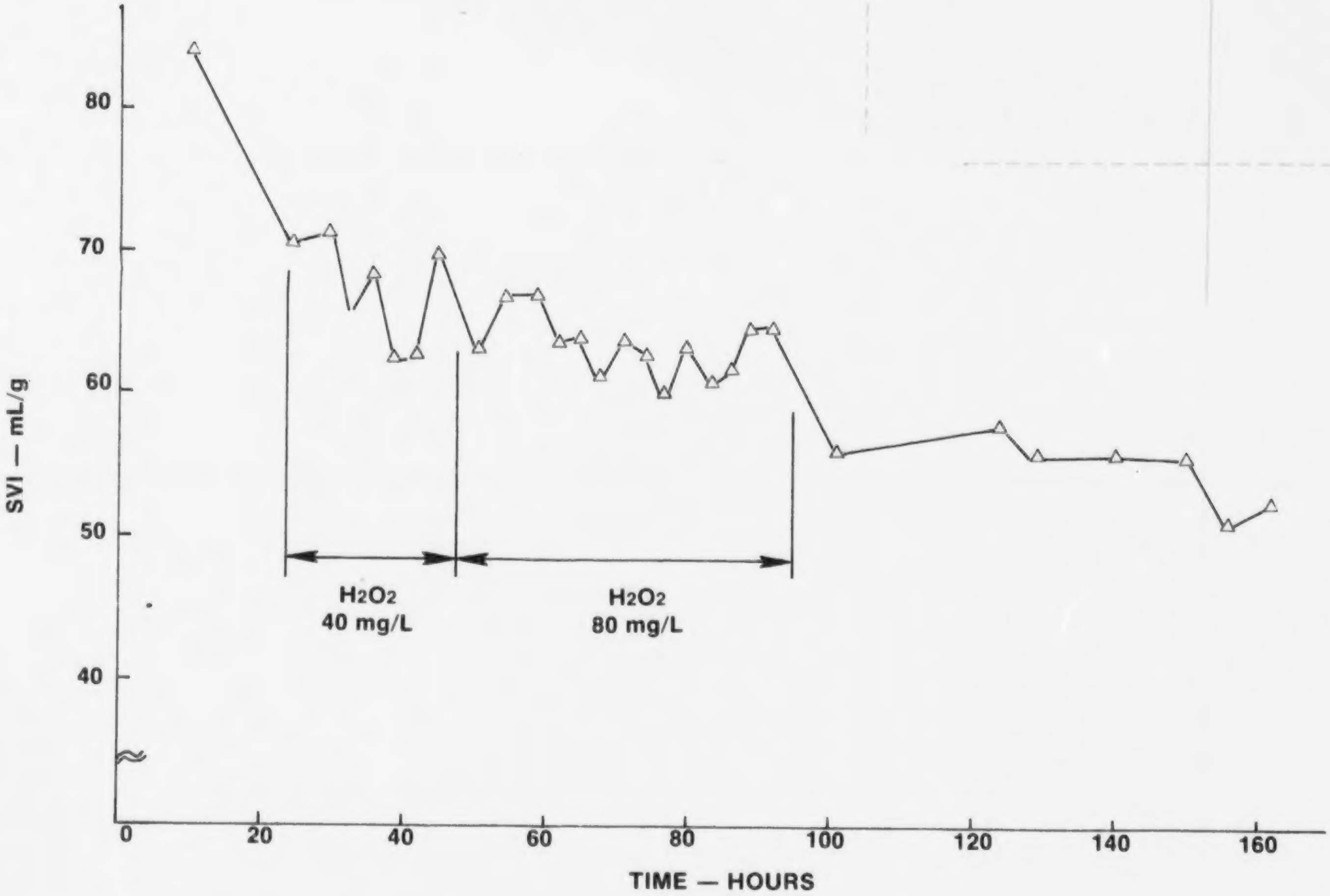

Figure C—32. Sludge volume index observed in full-scale H_2O_2 treatment.

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sulfides. Also, measuring the influent and effluent sulfide levels offered a control mode more objective than microscopic observation. The trial was started on October 26, 1977 and continued, in part, through November 22, Table C—47 shows the effect on operating parameters and filamentous growth of the H_2O_2 additive trial described below. the trial period is divided into groups of days and the parameters were expressed as averages during each period.

A 50 percent solution of H_2O_2 was fed directly into the primary clarifier influent line at 0.38 to 0.49 m^3/day (100 to 130 gal per day) for six days; at 0.19 m^3/day (50 gpd) for seven more days and at 0.11 m^3/day (30 gpd) seven days more. The H_2O_2 input was discontinued during the remaining six days of the trial period.

Referring to Table C—47, as SVI decreased and ISV and secondary clarifier underflow concentration increased, all three to plateau levels, the H_2O_2 addition rates could be dropped and still maintain the plateaus. At the same time both primary effluent and secondary effluent sulfide fell to zero or 0.1 ppm. Likewise, it was clear to microscopic exam that the filamentous population had fallen to 40 to 60 percent of the original.

The experiment was considered a success. In fact, even though H_2O_2 addition ceased in late November 1977, secondary underflow concentration stayed at 2.5 percent or higher until early September 1978. Other parameters, such as SVI and ISV, remained in normal ranges.

After an extended period of unexplained, high organic loads to the WWTP, filamentous growth accelerated and secondary underflow concentration dropped to 1.6 percent. Other parameters like MLSS, SVI, ISV, and effluent TSS were still within the normal range but, respectively were lower, higher, lower, and higher than the best values observed earlier. Anticipating worsening problems, H_2O_2 addition was started again on September 19, 1978 (50 percent solution) at a rate of 0.19 m^3/day (50 gpd) and stopped on November 13, 1978. The secondary underflow concentration rose to 2.4 percent and values of MLSS, SVI, ISV, and effluent TSS improved. Effluent residual sulfide during the addition period was zero to 0.1 mg/L. Filamentous population, by microscopic viewing was judged to be 50 percent lower. Again, the efficacy of H_2O_2 addition to the primary clarifier influent to oxidize the sulfides and control filamentous growth was proven.

TABLE C—47. EFFECT OF H_2O_2 INJECTION INTO PRIMARY CLARIFIER INFLUENT ON FILAMENTOUS ORGANISM GROWTH AND RELATED PARAMETERS

Parameter	0 days	1-3 days	4-6 days	7-12 days	13-19 days	20-24 days
H_2O_2 (50%), gpd	0	100-130	122-50	50	30	0
SVI, mL/g (avg)	72	58	48	46	48	—
ISV ft/hr (avg)	4.0	4.7	6.7	5.9	5.9	—
Sec underflow cond., % (avg)	1.7	2.0	1.8	2.2	2.0	2.7
Primary effluent sulfide, mg/L	—	—	0.3	0	0	0
Secondary effluent sulfide, mg/L	—	—	0.1	0.1-0	0	0
Filamentous population	—	No Change	—	40-60% reduction	—	—

Effluent control valve—The effluent level control valve performs a primary function of maintaining a certain level in an upstream foam pit. This serves to keep the effluent discharge line full which, in turn minimizes foam formation problems. Performing the primary function, the valve serves a secondary function of controlling the effluent discharge rate. Troubles with this valve have not been frequent but those which occurred caused serious delays or problems elsewhere in the WWTP system. The principal problems experienced were: (a) washout of valve housing structure before startup of WWTP, (b) failure of valve in closed position, (c) operation of valve in fixed open position, and (d) replacement of valve with "fail-in-open" position type.

The problems are described by letter excerpts and other pertinent discussions below. Each was reported to the SWCB as it occurred. In the case of problem "a"(114), we were running water through the discharge pipe from the secondary clarifiers to the diffuser in the river and found one leak in the horizontal section and another near a control valve housing which undermined the housing and opened a joint in the discharge line to the point it was not usable. Since then, we have installed pilings under the valve housing and taken other precautions to prevent another leak at that point.

This problem caused a one-month delay in WWTP startup.

Problem "b" had a "domino" effect on major parts of the WWTP. The problem and its mushrooming effects were described in a letter(115) to the SWCB.

This trouble began when a pin sheared on the control valve in the effluent discharge pipeline. When the pin sheared, the butterfly valve shut due to the force of the wastewater. As a result, the wastewater backed up and overflowed the effluent Parshall flume. The resulting overflow flowed into the secondary clarifier sludge pump pit, completely submerging the three sludge pumps and motors.

To prevent further flooding and to repair the control valve, the four mill lift station pumps were shut off. Also, the secondary sludge pumps were shut off. (This was done before the pumps and motors were submerged by the rising wastewater.)

When the valve pin was replaced, the valve was locked in the open position and the four mill sumps were started back up. However, secondary treatment was bypassed at the WWTP due to the totally submerged secondary sludge pumps (all flow went through primary treatment). The pump pit was then pumped empty with a portable gasoline sump pump and the pump motors were replaced with spare motors. About 8 hours later, one of the sludge pumps was started with approximately 70 percent of the wastewater flow directed to secondary treatment. After five more hours, a second sludge pump was started and the full flow was directed to secondary treatment.

To prevent this problem from recurring, we have modified the set point on the valve controller to minimize cycling of the valve.

In spite of the set-point adjustment, further examination of the valve indicated a distinct possibility of another failure. So, the valve was locked in the open position to avoid recurrence of the catastrophe described above. This was not the ideal solution as indicated in a letter(116) to the SWCB, which stated that we had locked the Wastewater Treatment Plant (WWTP) discharge foam control valve in the open position. This valve with its level controller, maintains a full discharge pipe (preventing entrainment of air) and prevents the discharge of foam from out WWTP. (With the valve in the open position, some foaming at our discharge occurs during low tide.)

Our instrumentation department inspected the valve and decided there was a danger that the valve would fail again. To prevent a re-occurrence of the trouble, we ordered a new counter-weighted butterfly valve that will fail in the open position.

As indicated above, the control valve, when operating properly, was effective in preventing foam. It became obvious that with the valve locked in the open position, foam problems would occur. This undesirable condition plus the danger of another "closed valve" failure led to the decision to replace the initial type valve with one which would, upon failure, be open, not closed.

Treated effluent diffuser—The treated effluent diffuser is described in detail in Appendix A. The diffuser installed differs only slightly from the diffuser design developed by Hydroscience, Inc.(43).

The effluent diffuser has operated well with the only one significant problem occurring not long after WWTP startup. The behavioral mode at that time indicated that a number of the nozzles must be plugged. This was a puzzling development until a professional diver discovered one-third the nozzles were, indeed, plugged with plastic film and a variety of other typical construction debris that had carried through the system, but could not pass through the 76 mm (3 in.) nozzles.

Since clearing the plugged nozzles in late September 1975, the effluent diffuser has operated trouble-free.

Sludge thickener—The sludge thickener is described in Appendix A. This equipment has not fulfilled its design function of sludge blending and thickening for a number of reasons. Initially, the rake mechanism was equipped with a single-speed drive that provided rake rotation at 0.042 rpm. It appeared that the rake arms were not moving the thickened sludge to the underflow pump suction as fast as it needed to be drawn away. A variable-speed drive was ordered August 30, 1974 and installed December 17, 1975. The top rotational speed of the rake with the new drive was 0.26 rpm.

Although the higher rotational speed was capable of moving thickened sludge to the pump suction, the higher rpm seemed to interfere with sludge thickening, thereby providing very little thickened sludge to be pumped.

An even more serious problem, which limited use of the thickener, was the septic, or anaerobic, condition that often developed at normal underflow rates and at the design detention time of 16 hours. Not only did the thickener contents emit gas and become odorous, the sludge underflow became more difficult to dewater. This, in turn, caused the vat level control at the dewatering drums to slow the flow from the thickener.

Without details, the piping, pumping, and valve system into and from the thickener was changed over a period of time to permit pipeline mixing of secondary and primary underflow and bypassing the thickener entirely. The system still permits use of the thickener under certain conditions mentioned below.

Changes were made in this system not only because of the septic events, but also because the primary and secondary underflows at averages of 4 and 2.0 to 2.5 percent, respectively, thickened no further in the thickener. When the secondary underflow drops below 2.0 percent, it can be diverted alone to the thickener, where the solids level does increase somewhat. However, most of the time now the thickener is not used at all for its intended purpose.

Sludge dewatering system—The equipment in the sludge dewatering system is described in Appendix A. There have been relatively few mechanical problems with this system. One 56 kW (75 hp) motor was replaced with a 75 kW (100 hp) motor on the filtrate pump to reduce overload frequency, hold or increase vacuum, and achieve higher cake solids as the primary/secondary sludge ratio went above the normal of 3 to 4/1 on a volume basis or 5 to 6/1 on a dry basis.

At one time, foam persistently developed in the vacuum receiver tanks, but this was corrected by installing fine-mist water nozles at the top of the tanks.

The design drum loading rate was 44 kg/hr/m^2 (9 lb/hr/ft^2); the actual loading rates have varied from 22 kg/hr/m^2 (4.5 lb/hr/ft^2) to 34 kg/hr/m^2 (7 lb/hr/ft^2) depending on the nature of the sludge mixture fed to the vats and the cake solids leaving the coil springs.

In general, the system design has proven adequate. There have been numerous times that cake solids off the filters was lower than desired. This increased the cost of, and more than doubled the difficulty in operating, a satisfactory landfill. However, in these cases feed-sludge quality was inferior because of an upset, or improper operating condition, elsewhere in the WWTP.

In another instance when the primary-to-secondary sludge ratio was exceptionally high and the quantity of primary sludge was also high, the filtrate pump motors overloaded and the cake formed was much lower in solids than normal. Again, this resulted from an unusually abnormal primary sludge situation.

Regardless of how satisfactory the dewatering system was compared with design expectations, the next section relates how important it is to increase cake solids. Higher cake solids came to be perceived as the panacea for easier sludge disposal by landfill or by any other method of disposal, such as incineration. Yet higher cake solids, with this type of sludge were not only elusive, they were frustratingly difficult to achieve in the range of 30 to 40 percent.

In fact, several types of pilot presses were tried(117) without much success in consistently reaching higher solids necessary to improve landfilling operations or to consider seriously the possibility of incineration. Chesapeake intends to continue more work to develop new methods, or to adapt an existing method, to moderate its problems in sludge disposal.

Sludge disposal system—Equipment used to haul and landfill waste sludge from the WWTP is described in Appendis A.

As explained in the two previous sections, primary and secondary clarifier underflows are mixed either in the blender-thickener or in a pipeline to the dewatering filters. The total amount of dry solids has been 10.9 to 21.8 metric tons/day (12 to 24 tons/day) at 12 to 22 percent solids concentration. The causes of such wide extremes are many and varied but involve all of the problems already mentioned in this report, such as black and green liquor spills to the WWTP, unusually low and high fiber losses from mill operations, and the level of filamentous organism population in the bacterial colony.

At solids concentrations of 12 to 15 percent, the sludge is essentially impossible to dispose of by landfill unless it is mixed with at least equal amounts of bark and/or dirt. Above 15 percent solids, disposal is easier but still requires the use of bark and dirt until the waste sludge alone reaches 19 to 20 percent solids. Under almost any circumstances, creating a landfill that is structurally stable, is free from potential leaching problems, and is insect and odor-free is very difficult. However, it was recently reported(118) that 90 percent of sludge disposal in the United States is by landfill. This is an impressive statistic and must indicate that in spite of landfilling difficulties it is the least expensive and most manageable method of sludge disposal.

Chesapeake had three areas approved by the Virginia Department of Health and the SWCB for sludge disposal. Two of the areas are small in size and would be filled quickly if used steadily; one is 9.7 km (6 miles) one way from the mill site and the other lies within 1.67 km (one mile). Both are viewed as emergency, or backup, disposal areas.

The third area, 14 miles one way from the mill site, is the largest site. It contains 131 ha (323 acres) with three fillable areas abutting. The terrain is rolling with a series of ravines and ridges. The first of these three contiguous areas was filled, earth-covered, and reseeded successfully. Following essentially the same procedures as used in the first area, the second area operation (involving a steeper and deeper ravine) proved to be a failure.

Building out from the ridge with the incoming loads of sludge, the bed reached a depth where it would not support the weight of the wide-track dozers used to distribute each load. Assuming some drying and stabilization would occur in a few days time, a new un-filled, adjoining section was started and built up in the same manner. This pattern was repeated several times until a section some 183 m (200 yards) wide had been filled. Frequent attempts were made to return to sections filled earlier but to no avail—they still would not support the equipment. During this time, the filled sludge underwent anaerobic decomposition, became odorous and more fluid, and became a haven for insects, principally flies. The ultimate occurred during a rainstorm—the beds of filled sludge collapsed, slid into the ravine and moved slowly, lava-like, toward a receiving stream. Chesapeake erected an earthen dam to halt the flow and to contain all leachate.

By this time, it was abundantly clear that this area must not only be just earth-covered; the entire area had to be mixed with dirt, then dirt covered over the entire slope, and the ravine itself partially filled. this area was then reseeded and abandoned for further filling. It was monitored closely for several months to assure that no odor or insect problems recurred.

Actually, for a relatively short period while covering the filled part of this second area, it was used for filling but by a different method. The sludge was spread into thinner layers and earth-covered daily. During this period, roadways were constructed and sandpits opened in the third area to prepare it for further filling.

In this third area, the thin-layer method has been continued. In addition bark and dirt have been mixed with the sludge, at times as much as 50 percent. Even so, operator diligence is a must to avoid layers that are too thick. Otherwise, the special wide-track dozers will sink into the mixture.

Without question, the present landfilling operation to dispose of sludge is far superior to early efforts in the two other fill areas. However, the frequency of low solids off the dewatering drums and the need to haul more bark and more dirt at these times leads Chesapeake to examine any other promising method of sludge disposal with serious interest.

Operating Performance Data For the WWTP—

The earliest compilation of operating data for the WWTP, especially compared with original design data, was presented by Djordjevic, et al(119) in early 1975. Table C—48, taken from the publication, shows the comparison between the original design and typical, initial operating data.

Regarding the data in Table C—48, Djordjevic and co-authors stated(119) that the most important differences between the expected performance and the actual performance at this point is the fact that the secondary sludge is not concentrating to the degree expected in the secondary clarifiers. This reduces the mixed liquor suspended solids level at a given recycle rate at well as increases the food to biomass loading at a given BOD_5 load to the plant. It is not clear at this point why the sludge is not settling and thickening as expected from the pilot studies. However, more filamentous growth appears to be present than that observed in the pilot plant studies. Also, the final effluent suspended solids levels are about twice as high as expected, although the final effluent BOD_5 is essentially the same. It is also not clear why the effluent solids are higher than expected or why the BOD_5 associated with these solids is lower than expected.

As noted previously, this reference to filamentous growth was probably a more accurate identification of the cause of the observed problems than realized. The data already shown in Table C—46 reaffirms the early observation. Certainly, the presence of filamentous organisms and the degassing of the biodegraded effluent in the secondary clarifier helps to explain the much higher effluent TSS. Why the BOD associated with the TSS was lower than that observed in the pilot plant work still is not understood.

Voluminous data have been collected on the WWTP for process control and for mandated reporting to the SWCB under NPDES Permit No. VA—0003115. The process control data are available from daily operating logs at Chesapeake. Data reported for permit purposes are on file at Chesapeake and at the SWCB.

Detailed data collected during a 33-day (May 18—June 19, 1977) evaluation of the WWTP were reported internally and are reproduced in Table C—49. Average operating characteristics of the WWTP, computed from the detailed data of Table C—49, are shown in Table C—50.

TABLE C—48. INITIAL OPERATING DATA FOR THE WASTEWATER TREATMENT PLANT

Item	Average Design	Operating
Influent UNOX		
Flow, L/s, (mgd)	13.1 (16.3)	529.5 (12.1)
BOD_5 (mg/L)	332	338
Suspended solids (mg/L)	85	88
Temperature (°F)	141	115—130
pH	8.6	8.5—9.5
UNOX Characteristics		
Reaction time based on flow (hr)	2.3	3.1
Mean sludge recycle (percent of flow)	28	35
Sludge recycle concentration, mg/L	25,000	16,700
Average mixed liquor suspended solids, mg/L	5,500	3,500
Average biomass loading F/M or F/M (kg BOD_5/kg MLVSS/day)	0.70	0.86
Average mixed liquor dissolved oxygen level (mg/L)	6	8
SVI	50	61
Overall oxygen utilization efficiency	90	89.4
Effluent UNOX		
BOD_5—total (mg/L)	27	30
BOD_5—soluble (mg/L)	10	9
Suspended solids (mg/L) (*)	50	100
Temperature, °C (°F)	35 (95)	36 (96)
pH	6.5	6.5—7.0

(*) Measured with Reeve Angel No. 934AH filter (sample size — 25 mL).

TABLE C—49. WWTP DATA FOR 33-DAY EVALUATION PERIOD
(MAY 18—JUNE 19, 1977)
PART I

		UNOX influent						
Date	Flow	BOD$_5$		TSS		Color		pH
1977	mgd	mg/L	lb/day	mg/L	lb/day	CU	lb/day	
5/18	11.66	563	54,749	93	9,044	2,220	215,883	6.68
5/19	11.36	530	50,213	77	7,295	2,411	228,424	6.74
5/20	11.84	470	46,410	93	9,183	2,100	207,366	7.05
5/21	11.83	467	46,075	88	8,682	2,067	203,935	7.18
5/22	12.12	454	45,891	94	9,502	2,488	251,489	6.95
5/23	12.21	390	39,714	215	21,894	1,684	171,484	6.95
5/24	11.63	355	34,433	80	7,760	1,608	155,967	7.93
5/25	10.69	445	39,674	77	6,865	1,723	153,613	9.02
5/26	9.99	455	37,909	57	4,749	1,570	130,807	9.13
5/27	12.06	518	52,101	80	8,046	2,641	265,633	8.05
5/28	12.30	525	53,856	61	6,258	2,794	286,614	7.77
5/29	12.53	413	43,159	71	7,420	1,378	144,001	7.19
5/30	13.23	385	42,480	77	8,496	1,761	194,306	7.00
5/31	13.25	378	41,771	110	12,156	2,067	228,414	7.04
6/1	11.47	395	37,786	79	7,557	1,914	183,093	
6/2	12.62	433	45,574	75	7,894	2,258	237,656	7.16
6/3	11.63	304	29,486	77	7,469	1,570	152,281	6.90
6/4	12.34	359	36,947	74	7616	1,914	196,980	6.82
6/5	12.84	318	34,053	78	8,353	2,297	245,976	7.00
6/6	12.86	390	41,828	93	9,974	2,220	238,100	6.97
6/7	12.41	420	43,470	79	8,176	2,258	233,702	713
6/8	13.20	515	56,695	102	11,229	2,144	236,029	6.77
6/9	10.30	374	32,127	97	8,332	1,684	144,659	6.70
6/10	12.12	427	43,162	98	9,906	1,455	147,073	6.75
6/11	12.66	355	37,482	94	9,925	1,684	177,804	6.96
6/12	12.89	407	43,754	77	8,278	2,373	255,104	6.92
6/13	11.30	305	28,744	72	6,785	1,416	133,477	7.45
6/14	11.94	548	54,570	142	14,140	3,177	316,364	8.59
6/15	12.72	458	48,587	159	16,867	2,526	267,970	7.15
6/16	11.88	455	45,081	126	12,484	2,373	235,115	7.10
6/17	12.75	438	46,575	116	12,335	2,297	244,252	7.00
6/18	11.78	422	41,459	102	10,021	1,990	195,508	6.81
6/19	9.59	367	29,353	323	25,843	2,029	162,281	7.00
Total	396.00	—	,405,168	—	330,525	—	6,841,330	—
Avg.	12.00	425	42,581	101	10,016	2,003	207,313	—

TABLE C—49. WWTP DATA FOR 33-DAY EVALUATION PERIOD (MAY 18—JUNE 19, 1977) PART II

	Secondary effluent						
Date	BOD$_5$		TSS		Color		pH
1977	mg/L	lb/day	mg/L	lb/day	CU	lb/day	
5/18	94	9,141	135	13,128	1,837	178,638	7.10
5/19	49	4,642	87	8,243	1,990	188,537	7.06
5/20	48	4,740	69	6,813	1,837	181,396	7.12
5/21	45	4,440	53	5,229	1,570	154,900	7.18
5/22	48	4,852	80	8,086	2,105	212,775	7.18
5/22	48	4,852	80	8,086	2,105	212,775	7.18
5/23	42	4,277	69	7,026	1,684	171,484	7.18
5/24	29	2,813	74	7,176	1,302	126,286	7.38
5/25	44	3,923	179	15,959	1,302	116,079	744
5/26	41	3,416	160	13,301	1,340	111,644	7.58
5/27	45	4,526	210	21,122	2,220	223,288	7.58
5/28	38	3,898	78	8,001	2,373	243,427	7.44
5/29	29	3,031	61	6,375	1,302	136,059	7.30
5/30	30	3,310	60	6,620	1,378	152,046	7.12
5/31	29	3,205	60	6,630	1,723	190,400	7.04
6/1	38	3,635	89	8,514	1,589	152,003	7.21
6/2	41	4,315	82	8,631	1,990	209,449	7.14
6/3	30	2,910	42	4,074	1,455	141,127	7.05
6/4	45	4,631	75	7,719	1,608	165,488	6.90
6/5	41	4,391	76	8,139	1,952	209,031	6.95
6/6	38	4,076	78	8,366	1,914	205,281	7.08
6/7	35	3,622	56	5,795	2,029	210,000	7.16
6/8	46	5,060	55	6,050	2,029	223,369	7.32
6/9	32	2,749	36	3,092	1,761	151,273	7.27
6/10	36	3,639	53	5,357	1,263	127,665	7.01
6/11	32	3,379	51	5,385	1,416	149,508	7.29
6/12	33	3,548	46	4,945	1,952	209,845	7.05
6/13	34	3,204	47	4,429	1,723	162,379	7.13
6/14	45	4,481	113	11,252	2,450	243,970	7.44
6/15	40	4,243	65	6,896	2,488	263,939	7.15
6/16	39	3,864	66	6,539	2,297	227,585	7.02
6/17	48	5,104	69	7,337	2,144	227,982	7.50
6/18	38	3,733	65	6,386	1,761	173,010	6.44
6/19	36	2,879	63	5,039	1,723	137,707	7.37
Total	—	133,677	—	257,366	—	5,977,570	—
Avg.	40	4,051	79	7,798	1,803	181,138	—

TABLE C—49. WWTP DATA FOR 33-DAY EVALUATION PERIOD (MAY 18—JUNE 19, 1977)

PART III

Date 1977	Removal efficiencies: BODR lb/day	BODR percent	ColorR lb/day	ColorR percent	Secondary underflow: Waste flow mgd	TSS mg/L	Daily avg lb/day
5/18	45,608	83.3	37,245	17.3	0.1236	14,386	14,832
5/19	45,571	90.8	39,887	17.5	0.1402	14,042	16,419
5/20	41,670	89.8	25,870	12.5	0.1429	15,347	18,290
5/21	41,635	90.4	49,035	24.0	0.1557	15,652	20,327
5/22	41,039	89.4	38,714	15.4	0.1812	17,305	26,149
5/23	35,437	89.2	0	0	0.0107	16,757	1,490
5/24	31,620	91.8	29,681	19.0	0.149	16,680	21,552
5/25	35,751	90.1	37,534	24.4	0.1549	16,680	21,552
5/25	35,751	90.1	37,534	24.4	0.0706	16,312	9,603
5/26	34,493	91.0	18,371	14.0	0.0394	15,923	5,235
5/27	47,575	91.3	42,345	15.9	0.0308	14,674	3,764
5/28	49,958	92.8	43,187	15.1	0.0797	16,062	10,680
5/29	40,128	93.0	7.942	5.5	0.1136	17,320	16,404
5/30	39,170	92.2	42,260	21.7	0.1292	16,692	17,979
5/31	38,566	92.3	38,014	16.6	0.1009	17,057	14,355
6/1	34,151	90.4	31,090	17.0	0.1149	16,650	15,958
6/2	41,259	90.5	28,207	11.9	0.1890	15,650	24,663
6/3	26,576	90.1	11,154	7.3	0.1474	15,868	19,511
6/4	32,316	87.5	31,492	16.0	0.1394	15,988	18,589
6/5	29,662	87.1	36.945	15.0	0.1886	15,800	24,855
6/6	37,752	90.3	32,819	13.8	0.1170	16,128	15,736
6/7	39,848	91.1	12,660	5.4	0.1296	19,368	20,926
6/9	29,378	91.4	-6,614	-4.6	0.4471	19507	72740
6/10	39,523	91.6	19,408	13.2	0.1633	17,938	24,436
6/11	34,103	91.0	28.296	15.9	0.1455	18,407	22,330
6/12	40,206	91.9	45,259	17.7	0.2666	18,010	40,043
6/13	25,540	88.9	-28,932	-21.7	0.1325	15,882	17,548
6/14	50,089	91.9	45,259	17.7	0.1104	15,948	14,684
6/15	44,344	91.3	4,031	1.5	0.1742	17993	26,147
6/16	41,217	91.4	7,530	3.2	0.0390	18,553	6,028
6/17	41,471	89.0	16,270	6.7	0.0661	18,140	9,997
6/18	37,726	91.0	22,498	11.5	0.0428	18,853	6,730
6/19	26,474	90.2	24,574	15.1	0.0726	16,218	9,816
Total	1,271,491	—	862,968	—	4.3323	—	614,163
Avg.	38,530	90.5	26,151	12.6	0.1313	16,769	18,611

TABLE C—49. WWTP DATA FOR 33-DAY EVALUATION PERIOD
(MAY 18—JUNE 19, 1977)
PART IV

	Mixed liquor			F/M
		TVSS under aeration		BOD/MLVSS
Date 1977	TSS mg/L	mg/L	lb	lb/lb
5/18	4,303	4,088	53,525	0.85
5/19	4,295	4,080	53,426	0.85
5/20	4,637	4,405	57,680	0.72
5/21	4,814	4,573	59,882	0.70
5/22	5,172	4,,913	64,335	0.64
5/23	5,042	4,790	62,718	0.57
5/24	4,635	4,403	57,659	0.55
5/25	5,150	4,893	64,066	0.56
5/26	5,408	5,138	67,276	0.51
6/27	4,634	4,402	57,647	0.83
5/28	4,862	4,619	60,483	0.83
5/29	5,230	4,969	65,061	0.68
5/30	4,603	4,373	57,261	0.83
5/28	4,862	4,619	60,483	0.83
5/29	5,230	4,988	65,310	0.52
5/30	4,603	4,373	57,261	0.68
5/31	4,907	4,662	61,043	0.63
6/1	5,250	4,988	65,310	0.52
6/2	4,880	4,636	60,707	0.68
6/3	4,645	44,13	57,784	0.46
6/4	4,474	4,202	55,022	0.54
6/6	4,026	3,825	50,083	0.15
6/7	4,030	3,829	50,133	0.79
6/8	4,420	4,199	54,985	0.54
6/8	4,420	4,199	54,985	0.54
6/9	4,777	4,538	59,426	0.49
6/10	4,460	4,237	55,482	0.71
6/11	4,365	4,147	54,301	0.63
6/12	4,177	3,968	51,962	0.77
6/13	4,498	4,273	55,955	0.46
6/14	4,542	4,315	56,502	0.89
6/15	4,355	4,137	54,176	0.82
6/16	4,820	4,579	59,961	0.69
6/17	4,967	4,719	61,789	0.67
6/18	5,290	5,026	65,808	0.57
6/19	4,898	4,653	60,931	0.43
Total	—	—	1,928,036	—
Avg.	4,697	4,462	58,425	0.66

TABLE C—49. WWTP DATA FOR 33-DAY EVALUATION PERIOD (MAY 18—JUNE 19, 1977) PART V

	Oxygen						
Date 177	Feed lb/day	Vent gas purity percent	Vent lb/day	Consumed lb/day	Utilization percent	O2 applied / BODR lb/lb	O2 Consumed / BODR lb/lb
5/18	46,884	38	4,755	42,129	89.9	1.03	0.92
5/19	54,498	39	4,461	50,037	91.8	1.20	1.10
5/20	50,963	38	3,644	47,319	92.8	1.22	1.14
5/21	49,791	37	4,531	45,260	90.9	1.20	1.09
5/22	53,094	38	3,987	49,107	92.5	1.29	1.20
5/23	50,476	37	4,241	46,235	91.6	1.42	1.30
5/24	48,370	37	3,274	45,096	93.2	1.53	1.43
5/25	45,802	41	1,647	44,155	96.4	1.28	1.24
5/26	41,317	52	453	40,864	98.9		
5/27	53,680	44	507	53,173	99.1	1.13	1.12
5/28	59,626	37	909	58,717	98.5	1.19	1.18
5/28	48,684	37	2,929	45,755	94.0	1.21	1.14
5/30	47,197	37	3,139	44,058	93.3	1.20	1.12
5/31	51,079	36	3,441	47,638	93.8	1.42	133
6/1	48,477	37	3,005	45,472	93,8	1.42	1.33
6/2	54,911	37	3,326	51,585	93.9	1.33	1.25
6/3	43,497	38	2,684	40,813	93.8	1.64	1.54
6/4	47,577	36	2,875	44,702	94.0	1.47	1.38
6/5	48,296	38	3,432	45,152	92.9	1.29	1.20
6/6	48,584	37	3,432	45,152	92.9	1.29	1.20
6/7	49,212	37	3,160	46,052	93.6	1.23	1.16
6/8	56,356	38	3,395	52,961	94.0	1.09	1.03
6/9	46,718	37	1,172	45,546	97.5	1.59	1.55
6/10	47,659	31	1,041	46,618	97.8	1.21	1.18
6/11	51,912	38	2,678	49,234	94.8	1.52	1.44
6/12	53,978	39	4,270	49,708	92.1	1.34	1.24
6/13	43,043	43	679	57,345	94.4	169	1.59
6/14	58,024	43	679	57345	98.8	1.16	1.14
6/15	57,231	44	1,147	56,084	98.0	1.29	1.26
6/16	57,958	37	4,008	53,950	93.1	1.41	1.31
6/17	61,898	38	4,261	57,637	93.1	1.49	1.39
6/18	59,139	37	4,064	55,075	93.1	1.57	1.46
6/19	53,283	37	4,273	49,010	92.0	2.01	1.85
Total	1,689,214	—	97,364	1,591,850	—	44.80	42.21
Avg.	51,188	38	2,950	48,238	94.2	1.36	1.28

The data in Tables C—49 through C—50 should be compared cautiously with pilot plant data. For example, the pilot plant work was done with no bleach plant effluent and with no significant filamentous growth in the mixed wastewater. The secondary underflow concentration in the pilot work ranged from 2.5 to 3.0 percent. During the 33-day full-scale WWTP evaluation, the bleach plant was operating at a production rate averaging 232 metric tons/day (256 tons/day) and filamentous organisms were present substantially in the mixed liquor. The latter can be deduced from the secondary underflow concentrations of less than 2.0 percent each day (see Table C—49) and the average value of 1.68 percent for the full test period. As already reported, when the filamentous population was reduced by H_2O_2 injection, full-scale secondary underflow concentration increased to 2.5 percent.

Recognizing but setting aside such caution momentarily, it would be interesting to examine whether or not full-scale and pilot results can be reasonably compared. Setting aside also for a moment the known, general effects of bleach plant effluent on WWTP performance and considering only the effect of filamentous growth, it appears possible to compare certain pilot and full-scale results. For example, referring to Table C—46, certain parameters do not appear to be strongly influenced by the presence of filamentous organisms. These are compared for the pilot plant results and for the 33-day full-scale evaluation in Table C—51.

TABLE C—50. AVERAGE OPERATING CHARACTERISTICS OF WWTP (FROM DETAILED DATA OF TABLE C—49)

Parameter	Operating Characteristics (avg)
Solids production kg TSS/kg BOD	0.43
Average biomass loading F/M or F/M_a or F/M (kg BOD_5/kg MLVSS/day)	0.66
Oxygen usage, kg applied/kg BOD	
Applied	1.36
Consumed	1.28
Oxygen utilization, percent	94.2
UNOX influent	
Suspended solids, kg/day (lb/day)	4,543 (10,016)
BOD_5, kg/day (lb/day)	19,315 (42,581)
Color, kg/day (lb/day)	94,037 (207,313)
Flow, L/s (mgd)	525 (12.0)
Secondary effluent	
BOD_5, kg/day (lb/day)	1,836 (4,050)
Color kg/day (lb/day)	82,164 (181, 138)
Suspended solids, kg/day (lb/day)	3,537 (7,798)
Removal efficiency	
BOD_5, percent	90.5
Color, percent	12.6

Computation of certain of the numbers in the "Pilot" column of Table C—51 involve concepts beyond the scope of this report. These concepts are explained more fully by Grader, et al(79).

With respect to pilot oxygen utilization in Table C—51, Grader et al(79) states on page 105 of the published article, "The oxygen utilization efficiency was consistently above 90 percent". Likewise, with respect to pilot BOD_5 removal efficiency the same published article stated that "Over this entire (F/M) range of 0.46 to 1.4, the oxygenation system was able to essentially maintain a constant total BOD_5 removal of 91 percent in the secondary system". The average total BOD_5 removal efficiency of 90.5 percent for the full-scale 33-day trial occurred over an F/M range of 0.43 to 0.94. Although, certain operating characteristics of the full-scale system fell short of the corresponding pilot system results, such as secondary underflow concentration, the agreement among parameters shown in Table C—51 is excellent.

Summarizing this section Wastewater Treatment System Evaluation, it is fair to say that the oxygen activated sludge system has served Chesapeake well in treating wastewaters from pulp and paper manufacturing. There have been numerous problems that caused upsets and some noncompliance situations under NPDES Permit No. VA—0003115. But solving these problems has been a series of learning experiencs which have beneficially corrected certain in-plant and WWTP procedures.

Although the full-scale system performance, insofar as the data can be compared, confirms the pilot plant results quite well, there is still much to be learned about the influence of bleach plant effluent on performance of the WWTP.

As flow of bleach effluent to the WWTP has started and stopped, temporary upsets, with typical symptoms, have been predictable. One outstanding question yet to be answered is the effect increased bleached pulp production and increased flow of bleach plant effluent on the performance of the WWTP. Recent incomplete evidence points toward another lengthy period of learning how to contend with more severe WWTP upsets caused by increasing bleaching effluent. Technically, the prospect, is challenging; from an operating and regulatory compliance standpoint, the prospect is dismaying.

TABLE C—51. COMPARISON BETWEEN CERTAIN PILOT AND FULL-SCALE OPERATING CHARACTERISTICS OF THE WWTP

Parameter	Operating characteristics Pilot	Full-scale
Solids production, kg TSS/kg BOD_r	0.39	0.43
Oxygen usage, kg O_2/kg BOD_r consumed	1.30	1.28
Oxygen utilization, percent	90+	94.2
UNOX influent BOD_5 kg/day (lb/day)	19,315 (42,581)	19,315 (42,581)
Secondary effluent BOD_5, kg/day (lb/day)	— —	1,838 (4,051)
Removal efficiency BOD_5, percent	91	90.5

Wastewater Treatment Plant Capital and Operating Costs—

The capital costs(78) of the WWTP and the in-plant colection system are shown in Table C—52. The capital costs of the cryogenic oxygen plant were reported previously.

Annual operating costs (1976) of the WWTP (including sludge dewatering) and an estimate of annual power costs only for the in-plant collection system are shown in Table C—53. Operating costs for the cryogenic oxygen plant were reported previously.

TABLE C—52. CAPITAL COSTS OF WASTEWATER TREATMENT PLANT AND IN-PLANT COLLECTION SYSTEM

System	Capital Cost*
WWTP	$3,200,000
In-plant collection	2,000,000

*1973 dollars

TABLE C—53. ANNUAL OPERATING COSTS OF WASTEWATER TREATMENT (AND ESTIMATE OF POWER COSTS FOR IN-PLANT COLLECTION SYSTEM)

Item	Annual Cost
Supplies and materials	
a. On-site oxygen at $20.35/ton	$222,939
b. Ammonia	33,500
c. Phosphoric acid	52,200
d. Defoamer	174,200
e. Polymer (sludge dewatering)	2,700
f. Gas, oil, and grease	2,200
g. Maintenance materials	108,300
h. Miscellaneous supplies and materials	10,200
Mechanical maintenance—labor	69,000
Operator labor (including fringe benefits)	216,700
Power at $21.9/1000 kWhr	172,125
Depreciation	399,200
Other	10,200
	$1,474,264
In-plant collection power costs 825 hp at $145/hr-year	120,000
TOTAL	$1,594,264

BLACK LIQUOR OXIDATION SYSTEM

Black liquor oxidation is a process intended to reduce sulfide gas emissions from kraft recovery systems. Basically, there are four types of BLO systems: (1) weak liquor with air, (2) weak liquor with oxygen, (3) heavy liquor with air, and (4) heavy liquor with oxygen.

System Selection

Chesapeake conducted trials starting in the Spring of 1976 to determine the feasibility of using tonnage oxygen to oxidize black liquor as an alternative to a conventional air oxidation system. The trials were made on three scales as follows:

(1) Bench-scale batch trials
(2) Large-scale mill trials
(3) Small-scale pilot trials

These trials established oxidation reaction rates and oxygen requirements to develop the economic comparison between oxygen and conventional air BLO system (see Appendix C) and subsequently determined the design criteria for the BLO system.

Laboratory Studies Using Oxygen—

The bench scale laboratory trials were run batchwise using a blender with a sparge for oxygen and air addition as stated in J. R. Smith's report dated February 15, 1971.(120) The batch trials were carried out to determine (1) the black liquor oxidation rate and (2) the required oxygen to sodium sulfide ratio. Knowing these factors, the operating and capital costs of air vs. oxygen systems were calculated.

The reaction rate was found to depend on the degree of mixing of the oxygen with the liquors. Reaction times as low as 14 seconds were measured. The O_2/Na_2S ratio required to oxidize black liquor was found to be about 0.53 kg O_2/kg Na_2S in the 88 to 93° C (190 to 200° F) temperature range. The ratio required to oxidize a prepared Na_2S solution was 1.19 kg O_2/kg Na_2S. The fact that a lower oxygen to sodium sulfide ratio occurred when the black liquor was oxidized led Smith to the conclusion that a "reaction causing autoxidation of the liquor by the sulfate . . . presents the possibility of oxidizing black liquor using less than the theoretical oxygen". He stated that "with adequate control, BLO might be made to operate using 0.41 kg O_2 per kg Na_2S, perhaps less."(120)

Mill-Scale and Small-Scale Pilot Trials—

Following the laboratory batch studies, questions still needed to be answered on (1) process economics, (2) reactor design and control, and (3) side effects. To answer these questions, a mill-scale trial was conducted from April 24 to May 12, 1972 using oxygen for black liquor oxidation of the A-set evaporator feed liquor. Further questions on reactor design for optimum economic operation of the process led to small-scale pilot trials in which the reactor dimensions could be easily changed. Both the mill-scale and small-scale pilot trials are explained in a report by J. R. Smith.(27).

As a result of the mill-scale and small-scale pilot trials, the following conclusions were reported on weak liquor BLO:

(1) A single-stage pipeline reactor could remove 95+ percent of the Na_2S with an oxygen utilization efficiency approaching 100 percent with a shear rate of at least 1.5×10^6 min^{-1} and a 30-second retention time in the reactor.

(2) A sparge lance with 1.6 mm (1/16-inch) holes will adequately disperse oxygen in the reactor.

(3) The reactor pressure should be about 276 kPa (40 psig) or less.

(4) Pipe friction losses will be increased with oxygen addition.

(5) The efficiency of oxygen usage will depend on the efficiency of oxygen control.

(6) The residual Na_2S concentration can be controlled to +0.24 g/L in the range of 0.5 to 1.0 g/L Na_2S.

(7) No reversion back to Na_2S in heavy liquor storage is expected.

(8) A temperature increase of about 7° C (12° F) in the BLO reactor is expected.

(9) A 10 percent increase in tall oil recovery is expected with BLO.

(10) There was less odor in evaporator condensates with BLO.

(11) BLO had no adverse effects on evaporator operation.

Small-scale pilot trials were also run on heavy black liquor using both air and oxygen to develop some design criteria for heavy liquor polishing.(27) The results of the trials were as follows:

(1) Air heavy liquor polishing required 150 percent theoretical air and produced a viscous, foamy liquor with a specific gravity of 0.6.

(2) Oxygen heavy liquor polishing (operated with the following parameters:

  (a) The O_2/Na_2S ratio was 0.53 kg O_2/kg Na_2S or an oxygen utilization efficiency of 77 percent.

  (b) The reactor residence time required to achieve 0.95 g/L residual Na_2S was 43.4 seconds and the time required to achieve 0.50 g/L residual Na_2S is estimated at 67.7 seconds.

  (c) Reactor shear rate was 21,650 min^{-1} and the Reynolds number was 1500.

It was decided that action on heavy black liquor polishing would be taken at a later date.

Process Design, Engineering and Installation

The work done on the BLO pilot trials led to the design of the full-scale BLO system based on the following considerations:

(1) Shear rate
(2) Residence time
(3) Pressure
(4) Temperature
(5) Method of oxygen addition
(6) Process control scheme
(7) Electrode selection

Shear Rate—

The pilot trials showed that the reaction rate and reaction efficiency were more accurately predicted by shear rate than by Reynolds number which has been used as the basis for design on other BLO systems. Using shear rate for predicting reactor efficiency was supported by the following hypothesis: The oxidation of the sulfide has a considerably more rapid kinetic rate than the oxidation of thiosulfate or the oxidation of lignin, such that the sulfide oxidation is diffusion-controlled whereas the other oxidations are rate-controlled.(27) Therefore, the correlation of O_2/Na_2S versus shear rate allowed the scale-up of the pilot-scale BLO system for rapid Na_2S removal while maintaining a high oxygen utilization efficiency. The design shear rate of the full-scale system was 1.5×10^6 min.-1 which based on this correlation, would result in an oxygen utilization efficiency approaching 100 percent (0.41 kg O_2/kg Na_2S). This predicted oxygen to sulfide ratio was used to determine the amount of oxygen required for the system.

Residence Time—

Chesapeake experiments indicated that under normal conditions of weak liquor Na_2S concentrations, at least 30 seconds is required for 99+ percent reaction of the sulfide.(27) Underdesigning for residence time may result in waste of oxygen, i.e., by gassing off unreacted oxygen in the receiving vessel. Longer residence time than necessary is of no benefit. In the mill scale trial, almost all of the oxygen was consumed in 10 seconds, which suggested that 10 seconds was a practical minimum requirement for the amount of time for exposure to high shear. Therefore, the full-scale system was designed with a total residence time of 45 seconds, with 10 seconds of this time at the design shear rate. The existing evaporator feed line allowed the desired residence time, and the 10 seconds at the design shear rate was provided by replacing 30.5 m (100 ft) of 254 mm (10-in.) diameter pipe with 203 mm (8-in.) diameter pipe.

Pressure—

The BLO pilot(27) trials showed that operating the reactor at pressures much higher than 276 kPa (40 psig) may cause some loss in oxgen usage efficiency. Some piping changes were made in the evaporator feed lines to allow a lower weak liquor pressure at the header feeding the evaporators, hence a lower reactor pressure. The design pressure for the weak liquor entering the BLO reactor was 448 kPa (65 psig), with an oxygen injection pressure of 482 kPa (70 psig).

Temperature—

Temperature can affect efficiency insofar as the end product of the sulfide oxidation can be either thiosulfate if the reaction is carried out at above 71 ° C (160° F) or elemental sulfur is carried out below 71° C (160° F). The sulfur-producing reaction is more efficient in that it requires less theoretical oxygen, but it is an undesirable reaction because the sulfur reverts back to sulfide during multiple-effect evaporation. Hence, the Chesapeake weak liquor temperature of 88° C (190+ ° F) is satisfactory, as borne out by the experimental results. A temperature rise of about 7° C (12° F) will result from the oxygen addition, which will increase evaporator capacity slightly due to the temperature increase.

Method of Oxygen Addition—

Mill trials showed that oxygen could be injected and dispersed adequately in a single stage reactor using a sparge lance with 1.6 mm 1/16-in. holes.(27) The BLO system was designed with two sparge sections to provide an alternate in the event of pluggage problems or maintenance required on one of the sparge sections. Originally, each of the two sparge sections were designed with two 38 mm diameter (1½-in.) stainless steel sparge pipes arranged one over the other, approximately 172 mm (5-in.) center to center, in a 254 mm diameter (10-in.) carbon steel tee section of liquor piping. Each sparge pipe had 184 holes, 3.2 mm in. diameter (⅛-in), arranged in 4 rows of 46 holes each.

Process Control Scheme—

The efficiency of the oxygen usage for BLO depends on the oxygen control. the oxygen is controlled by measuring the residual Na_2S and sending a signal to an analyzer controller which adjusts the setpoint on a ratio controller. This ratio controller receives an input signal of the black liquor flow which in turn adjusts the setpoint on an oxygen controller according to the ratio set by the analyzer controller. The oxygen controller adjusts the oxygen flow based on the signal it receives from the ratio controller (Appendix A, Figure A—14).

Electrode Selection—
During the BLO trials,(27) the black liquor Na_2S concentration was measured with a silver billet and calomel reference electrode pair. However, considering the recent advancements made in specific ion electrodes, it was decided to try using a sulfide specific ion electrode for measuring the Na_2S concentration of the oxidized weak black liquor.

Tests using the sulfide specific ion electrode showed no correlation between the electrode output and the Na_2S concentration. It was later determined that the sulfide ion electrode was designed to measure the concentration of free $S^=$ ions only. Also, during the tests, it was determined that the calomel reference electrode was not suitable for operation under pressure. Therefore, it was decided to go back to the silver billet electrode used in the pilot trials and use a Tamal Bellomatic (thallium/KCl) reference electrode supplied by Electrofact that could be used under pressure.

Tests using the silver billet/thallium electrode pair gave a good correlation of output versus Na_2S concentration. Silver bilet electrodes manufactured by Electrofact, Bechman, and Leeds & Northrup were evaluated. Very little difference was found between the various silver billet electrodes. It was decided to use the Electrofact silver billet electrode since we were using their reference electrode already.

BLO System Evaluation

Mechanical, Process, and Control Problems—
Following the start-up of the BLO system, several problems occurred in the operation of the process. Some of the problems and their solutions are reported below.

Compressor seal—The oxygen compressor is a Nash H-6 compressor with a 149 kW (200 hp) motor designed to deliver 13.6 m^3/min (480 cfm) of oxygen at inlet conditions of 227 kPa and 21° C (33 psig and 70° F) with a discharge pressure of 33.8 kPa (70 psig). The compressor requires about 1.9 L/s (30 gpm) of seal water and incorporates a centrifugal separator to separate the water from the oxygen. The level control valve on the separator has created some operating problems on the BLO system. The bronze float valve failed to open far enough in some cases, causing the water to be carried downstream with the oxygen. Also, after a few months operation, the valve seat (bronze) became so badly eroded that the valve would not retain enough water in the separator resulting in oxygen blowing through the separator valve and out the seal water drain pipe. These two problems were remedied by changing the entire float valve assembly to stainless steel.

Oxygen pipeline—After start-up of the BLO system, it was discovered that the sparge system was sufficiently oversized that only one of the two sparge pipes needed to be used for oxygen injection. Later, it was suspected that the oxygen utilization efficiency, as well as the system operation, could be improved by modifying the sparger in the following ways: use a sparger with 1.6 mm diameter (1/16 in.) holes instead of 3.2 mm diameter (⅛-in.) holes to take advantage of better oxygen dispersion and more pressure drop across the sparger to aid system operation; and change the black liquor piping in the sparger section from 254 mm diameter (10-in.) to 203 mm diameter (8-in.). These two changes increased the oxygen utilization efficiency by about 1 percent and resulted in better operation of the BLO system. Figure A—15 (Appendix A) shows the sparger section after modification.

Oxygen flow control—The oxygen flow was originally controlled by a 762 mm (3-in.) fishtail butterfly valve just a few feet downstream from the oxygen flow orifice. Downstream some 152 m (500 ft) just prior to the oxygen spargers, is a safety shut-off valve which is designed to shut-off the oxygen to the spargers should the oxygen pressure fall to within 13.8 kPa (2 psig) of the black liquor pressure to prevent liquor from backing into the oxygen line. During start-up, with low oxygen flow, the flow control valve was reducing the oxygen line pressure sufficiently that the safety valve was closing because of the low oxygen pressure. With the safety valve closed, the oxygen line pressure could build up until it satisfied the differential pressure requirement and the valve reopened. This resulted in the safety valve cycling open and closed about every three seconds, causing very unstable operation. Since the safety valve was downstream of the pressure tap from the oxygen line to the differential pressure transmitter, the problem was solved by making the safety valve the oxygen flow control valve also. The safety valve, however, was a 102 mm (4-in.) valve; therefore, it was replaced with the original 76 mm (3-in.) flow control valve to afford better oxygen flow control.

Weak black liquor flow and pressure control—When the pressure control loop for the black liquor was designed, the pumps and pressure control valve were oversized. The pump was producing excessive head and the pressure control valve was operating in an almost closed position to keep the liquor pressure low enough for oxygen addition. The pressure control valve even went all the way closed on some occasions, which stopped all liquor flow to the evaporators. Since the black liquor pressure was higher than necessary, the impeller on the black liquor pump was trimmed, resulting in a lower black liquor pressure as well as a significant reduction in power consumption. The black liquor pressure control valve was also changed from an in-line valve to a valve in a recirculation line back to the weak black liquor storage tank for better control and more flexibility over a wider range of flow.

Steps were also taken to correct for the variation in pressure drop across the reactor due to normal liquor and oxygen flow changes. Originally, the black liquor pressure was controlled just prior to the sparger, thus a change in oxygen or liquor flow would result in a change in the black liquor feed pressure to the evaporators. In order to maintain constant black liquor feed pressure to the evaporators, the black liquor pressure was controlled at the header where the liquor flow splits to go to each set of evaporators. This resulted in much easier evaporator operation.

Sulfide electrode—Before the start-up of the BLO system, several different types and manufacturers of electrode pairs for sodium sulfide measurement were tested. It was decided to use a silver tip electrode and a Tamal Bellomatic reference electrode supplied by Electrofact, which can be used under pressure, for this application. It was discovered later that the life of these electrodes could be tripled by using heat shrink tubing to protect the electrodes from attack by the hot caustic liquor.(121)

Control system corrosion—A large amount of the system downtime has been caused by instrumentation malfunction. The instrument panel is located in the evaporator room and the environment has been detrimental to the electronic equipment (Fisher AC^2). therefore, to try to eliminate this problem, the instrument panel was sealed and air purged. Problems with the electronic instrumentation has been almost eliminated since this change.

BLO Oxygen Utilization Efficiency—

Installed system measurements—Measurements were made on many of the process variables on the full-scale BLO system. The operating data collected, and the normal value of each is reported in Table C—54.

TABLE C—54. FULL-SCALE BLO OPERATING DATA

Item	Before 1-78	After 1-78
Unoxidized B.L.		
Solids, percent	15	14
Temperature, °C (°F)	96 (204)	96 (204)
Na_2S conc. (g/L)	7.5	7.0
Oxidized B.L.		
Solids, percent	15.5	14.5
Temperature, °C (°F)	108 (226)	107 (225)
Na_2S conc. (g/L)	0.1	0.1
B.L. from evaporator		
Solids, percent	50	50
Na_2S conc. (g/L)	0.99	0.95
B.L. from H.L. storage		
Solids, percent	50	50
Na_2S conc. (g/L)	1.58	1.50
Flows		
B.L. flow, L/s (gpm)	122.8 (1950)	138.6 (2200)
O_2 flow kg/L (lb/hr)	1724 (3800)	1814 (4000)
Reactor shear rate (min^{-1})	-1.98×10^6	-2.45×10^6
Oxygen utilization efficiency, percent	76	77

From the data that was collected the following conclusions were reached:

(1) The solids content of the weak black liquor increases by about 0.5 percent with BLO.

(2) The exothermic reaction of BLO raises the black liquor temperatures by about 1.7° C (3° F) per gram per liter of Na_2S oxidized.

(3) The elevated temperature of the weak liquor feed results in an increase in evaporator capacity of about 2 to 3 percent as reported by the operators.

(4) There is a 2 percent reversion of the oxidized black liquor back to Na_2S on a weak liquor basis as the liquor goes through the evaporators, plus another 2 percent reversion while the liquor is in the heavy liquor storage tank.

(5) The pressure drop across the BLO reactor was increased by about 3.05 to 3.66 m (10 to 12 ft) of water head with the addition of oxygen.

In an attempt to determine the factors that most influence the oxygen utilization efficiency (OUE), regression analyses were run on several of the process variables vs. the oxygen utilization efficiency. A summary of the regression analyses for the full-scale BLO system is given in Table C—55 below:

TABLE C—55. REGRESSION ANALYSES OF FULL-SCALE BLO DATA

Equations*	R-squared
EFF = 2.715 NS + 53.58	0.3085
EFF = 0.00876 BLF + 60.62	0.2057
EFF = 5.745 RS + 66.47	0.1906
EFF = 1.438 OS + 74.06	0.0350
EFF = 4.446 RS + 2.485 NS + 48.6	0.4354
EFF = 0.00661 BLF + 2.323 NS + 45.62	0.4191

*Based on data collected 104/74 — 1/31/75

Where:

EFF = Oxygen utilization efficiency (OUE)
NS = Inlet Na_2S concentration, g/L
BLF = Black liquor flow gpm
RS = Shear rate x 10^{-6}, in.$^{-1}$
OS = Outlet Na_2S concentration, g/L
R-squared = Square of the correlation coefficient

Of the four variables measured (NS, BLF, RS, and OS), the inlet sulfide was found to have the highest correlation coefficient. The outlet or residual sulfide concentration had essentially no correlation with OUE. Multi-variable regression analyses resulted in a higher correlation coefficient than the single-variable analyses. The shear rate (RS) and inlet sulfide (NS) most accurately predicted the OUE with an R-squared of 0.4354. Since the BLO system must operate with an inlet Na^2S concentration set by other pulp mill criteria, e.g., white liquor sulfidity or washer dilution, shear rate is the only variable that can be manipulated to optimize the OUE.

The shear rate of the full scale BLO system cannot be easily changed; therefore, it was decided to extend the pilot trials to further study the effect of shear rate on OUE and also confirm the correlation coefficient of the other variables with OUE.

Parallel pilot system measurements—The parallel pilot trials were run from April 22 to July 26, 1976. The trials began using 36.6 m (120 ft) of 12.7-mm diameter (½-in.) hose with sample points at 9.15, 18.3, 27.5, 30.5, 33.6, and 36.6 m (30, 60, 90, 100, 110, and 120 ft) down the reactor as shown in Figure C—33. It was found that the reaction had gone to completion at the second sample point (18.3 m [60 ft] down the reactor), which corresponds to a retention time of about 7 seconds with the maximum flow using this pilot arrangement of about 0.33 L/s (5.2 gpm). Therefore, the pilot reactor was shortened to 18.3 m (60 ft) of 12.7 mm diameter (½ in.) hose with sample points at 9.15, 12.2, 15.3, and 18.3 m (30, 40, 50, and 60 ft) to allow higher reactor shear rates. Even with the higher flow rates afforded by the shorter reactor (about 0.44 L/s [7 gpm]), the reaction was completed by the 12.2 m (40 ft) sample point—a retention time of only 3.5 seconds.

After gaining some experience in operating the parallel pilot system, it was decided to operate the pilot system at four distinct shear rates for evaluating the effect of each of the variables evaluated on the full-scale system on the OUE. The four shear rates were as follows: (1) -3×10^6 min^{-1}; (2) -2×10^6 min^{-1}; (3) -1×10^6 min^{-1}; and (4) -0.2×10^6 min^{-1}. We attempted to hold the temperature, pressure, and outlet Na_2S concentration constant during the trials.

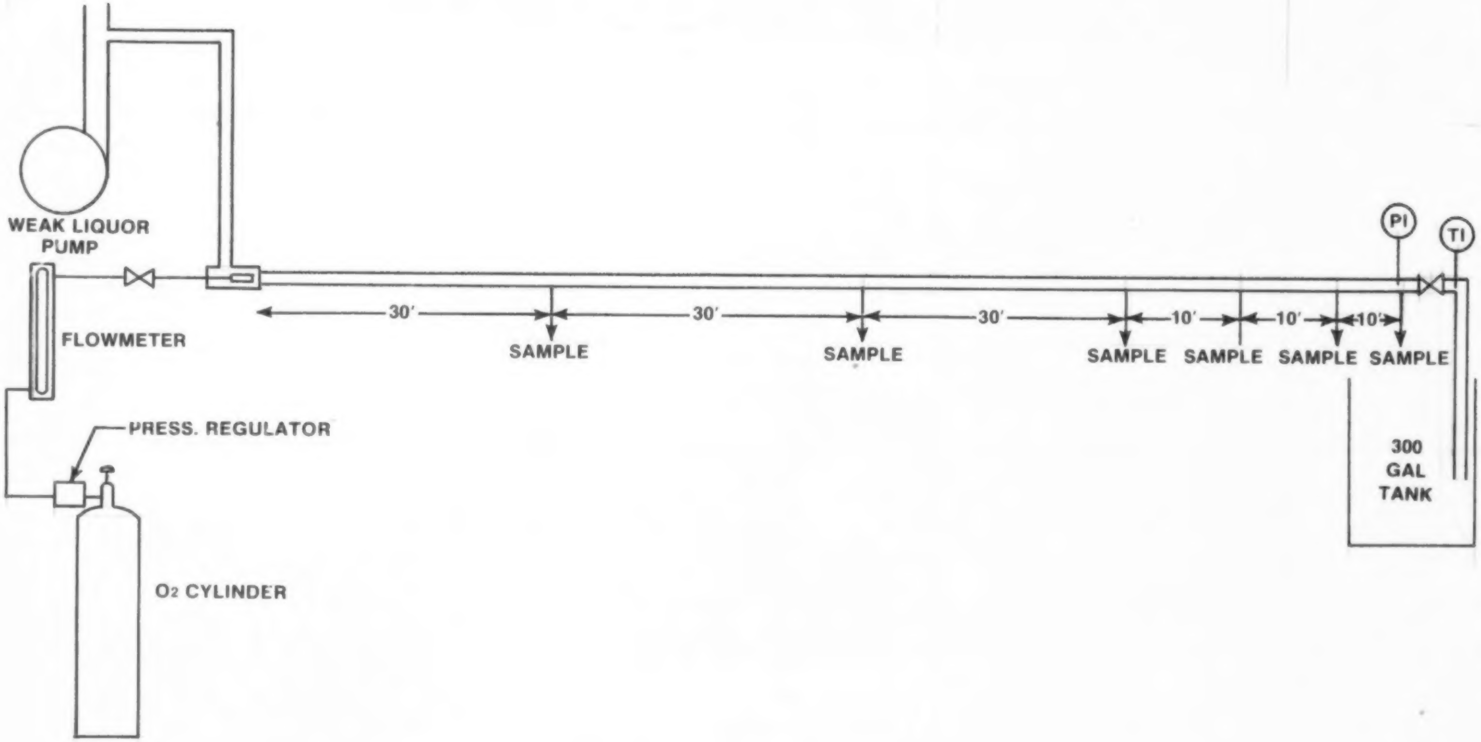

Figure C—33. Pilot BLO reactor.

A summary of the data collected in the parallel pilot system is reported in Table C—56. A regression analysis was run on each of the variables in Table C—56 to determine its correlation with OUE.

Multi-variable regression analyses were also run on black liquor flow and inlet sulfide concentration vs. OUE as was done on the full-scale BLO system. The results of the regression analyses are reported in Table C—57 following.

The R-squared values from the regression analysis of each variable as well as for the mulit-variable analyses are very close to the R-squared values determined in the evaluation of the full-scale data. The inlet Na_2S concentration versus OUE resulted in the highest R-squared value of the single-variable analyses as with the full-scale data. Also the multi-variable R-squared values were higher than any single-variable R-squared as with the full scale data.

From the data that was collected on the full-scale and pilot-scale, it was concluded that the only feasible way to increase the oxygen utilization efficiency was to increase the shear rate. Therefore, we

TABLE C—56. PARALLEL PILOT-SYSTEM DATA 5/28/76—7/20/76

RS Shear rate (min^{-1})	BLF B.L. flow, L/s (gpm)	EFF % EFF	NS Inlet Na_2S (g/L)	OS Outlet Na_2S (g/L)
-3.50×10^6	0.448 (7.11)	63.7	5.86	1.02
-3.37×10^6	0.435 (6.91)	65.1	6.24	1.24
-2.97×10^6	0.408 (6.47)	59.9	6.40	0.39
-2.90×10^6	0.402 (6.38)	59.6	6.94	0.60
-2.95×10^6	0.406 (6.44)	67.9	7.26	0.26
-2.98×10^6	0.408 (6.48)	56.3	5.54	0.24
-2.19×10^6	0.342 (5.43)	64.7	5.30	1.24
-2.39×10^6	0.360 (5.72)	68.3	5.46	1.50
-2.07×10^6	0.331 (5.26)	58.2	5.22	1.66
-2.25×10^6	0.348 (5.52)	62.5	5.86	0.34
-2.12×10^6	0.336 (5.33)	56.4	5.54	0.54
-2.12×10^6	0.336 (5.34)	53.5	5.08	0.24
-2.30×10^6	0.352 (5.59)	70.1	6.36	0.70
-1.19×10^6	0.241 (3.83)	56.8	5.38	0.46
-1.04×10^6	0.224 (3.56)	62.7	6.24	0.42
-1.04×10^6	0.224 (3.56)	57.4	5.70	0.46
-1.02×10^6	0.222 (3.52)	60.7	4.84	0.38
-1.01×10^6	0.220 (3.50	58.6	4.84	0.48
-0.96×10^6	0.214 (3.39)	53.9	4.68	0.20
-0.21×10^6	0.089 (1.41)	60.7	4.56	0.40
-0.22×10^6	0.093 (1.47)	57.7	4.18	0.18
-0.14×10^6	0.072 (1.14)	53.0	3.72	0.34
-0.23×10^6	0.095 (1.51)	60.9	6.24	0.28
-0.20×10^6	0.088 (1.39)	56.3	7.68	0.11
-0.17×10^6	0.081 (1.28)	60.8	6.74	0.76
-0.24×10^6	0.098 (1.55)	54.4	3.71	0.20
-0.11×10^6	0.061 (0.965)	41.5	3.12	0.54

decided to conduct pilot trials using a high shear in-line mixer manufactured by Kenics Corporation. During these trials we wanted to keep the black liquor flow rate the same with and without the Kenics mixer. We also wanted to maintain constant liquor feed pressure, temperature, and outlet Na_2S concentration as in the parallel pilot system.

The maximum flow through the 18.3-m (60-ft) reactor decreased from 0.44 L/s (7 gpm) without the mixer to about 0.22 L/s (3.5 gpm) with the mixer because of the increased pressure drop with the mixer. Therefore, the flow has to be throttled when the Kenics mixer was not in use to maintain the same flow rate.

Based on the four pairs of trials run with and without the Kenics mixer reported in Table C—58 it was concluded that there was no improvement in the oxygen utilization efficiency using the mixer. Because of the high pressure drop experienced with the mixer and since there was no indication of increasing the OUE with a mixer the trials using an in-line mixer were suspended.

TABLE C—57. REGRESSION ANALYSES OF PARALLEL PILOT-SYSTEM DATA

Equations*	R-squared
EFF = 3.163 NS + 41.9	0.3624
EFF = 1.438 BLF + 53.46	0.2740
EFF = 2.555 RS + 55.35	0.2588
EFF = 5.489 OS + 56.23	0.1561
EFF = 1.512 RS + 2.462 NS + 43.41	0.4353
EFF = 0.8838 BLF + 2.424 NS + 42.37	0.4461

*Based on data collected 5/28/76 — 7/20/76

Where:

EFF = Oxygen utilization efficiency (OUE)
NS = Inlet Na_2S concentration, g/L
BLF = Black liquor flow, L/s (gpm)
RS = Shear rate x 10^6, min^{-1}
OS = Outlet Na_2S concentration, g/L

TABLE C—58. OXYGEN UTILIZATION EFFICIENCY FOR PARALLEL PILOT SYSTEM USING KENICS MIXER (%)

No. of trial pair	With mixer	Without mixer
1	54.1	61.0
2	59.9	60.6
3	61.2	63.4
4	65.7	57.7
Avg	60.2	60.7